

dESIGN OF DISTILLATION COLUMN CONTROL SYSTEMS

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Preface

this is a book about the design of distillation column control systems. It is written primarily from the standpoint of an engineering design organization, and is based on years of experience with large design projects as well as on personal plant experience. Most new investment dollars go into new or modernized facilities, and it is in the design phase of projects for these facilities that the most opportunities occur and flexibility exists to influence process control. Consequently this book is aimed primarily at design personnel. It is our hope, however, that it will also be useful to those who have to operate or troubleshoot existing plants.

Part I is an introduction, including a perspective on control and a brief review of fundamentals of distillation, with emphasis on topics that will be of interest to the control engineer rather than to the column design engineer. The distillation review, it is hoped, will be particularly useful to nonchemical engineers.

Part II of the book, on concepts and configurations, discusses some practical aspects of distillation control. Once the requirements for a particular column in a particular process are understood, design engineers must make at least a preliminary choice of equipment arrangements and control system configuration. In this section we have mostly avoided the use of mathematics and control theory. It is our hope that our discussions of equipment and control system arrangements will be useful to process engineers, production supervisors, maintenance engineers, and instrument engineers seeking guidelines, alternatives, and perspectives.

Part III focuses on the quantitative design of distillation control systems. It is aimed at professional control engineers and any others concerned with the numerical definition and specification of control system performance. Probably the most important development in process control system design since about 1950 was the evolution of a substantial body of theory and mathematics, plus a large catalog of control system studies. Together, these permit quantitative design of most process control systems with a considerable degree of multivariable control. It is the purpose of this book to indicate the range of this technology, which has been developed for distillation control, to the point where it can be economically and reliably used for design. The ultimate economic advantages include lower plant investment (particularly in tankage), lower operating costs, and closer control of product quality. For the most part, we have stayed with

the modest theory of single-input, single-output (SISO) systems presented in previous books: *Techniques of Process Control* by P. S. Buckley (Wiley, 1964) and *Process Modeling, Simulation, and Control for Chemical Engineers* by W. L. Luyben (McGraw-Hill, 1973). This kind of theory and mathematics not only is adequate for noninteracting systems and for simple interacting systems, but it has the advantages of requiring minimum formal training and of permitting low design costs. "Modern" or "optimal" control techniques are mentioned only briefly here because their use on real, industrial-scale distillation columns has been quite limited to date. These techniques are still being actively researched by a number of workers, and it is hoped that they eventually will be developed into practical design methods. As of the date of the writing of this book, however, these mathematically elegant methods are little used in industry because of their complexity, high engineering cost, and limitation to relatively low-order systems. Simulation techniques also are not covered since there are several texts that treat this topic extensively.

In the past five years, we have witnessed the introduction and proliferation of microprocessor-based digital controls of various sorts that are intended to replace analog controls. In fact, most of the newly installed control systems are of this type. In addition, we are seeing more control being implemented in process control computers. Sampled-data control theory has taken on new importance because of these developments and so we have included a chapter on previous work we have done in this area as it relates specifically to distillation columns. The concepts we present are quite basic as opposed to the recent advances in adaptive, multivariable, and predictive control, but we hope they will benefit those interested in synthesizing single-loop sampled-data controllers.

Many thanks are due our associates in the Du Pont Company, particularly R. K. Cox, and throughout the industrial and academic communities for helpful comments and suggestions. Many of the concepts presented in this book have been vigorously debated (over untold cans of beer) during the Distillation Control Short Courses held at Lehigh University every other spring since 1968.

We also wish to thank Leigh Kelleher for major assistance in formatting and editing, Arlene Little and Elaine Camper for typing, and Ned Beard and his Art Group for preparing the illustrations.

Page S. Buckley
William L. Luyben
Joseph P. Shunta

Nomenclature

In this work an effort has been made: (1) to use symbols and units commonly employed by chemical engineers, (2) to define each symbol in a chapter when the need for that symbol arises, and (3) to keep symbols and units as consistent as possible from chapter to chapter. A few symbols, however, have different meanings in different parts of the text. The list that follows contains the major symbols and their usual meanings:

<i>a</i>	transportation lag or dead time, usually seconds or minutes
<i>A</i>	area, ft ²
<i>B</i>	bottom-product flow, mols/min
<i>c</i>	specific heat, pcu/lbm °C
<i>C</i>	acoustic capacitance, ft ⁵ /lbf
<i>C_v</i>	control-valve flow coefficient, gallons per minute of water flow when valve pressure drop is 1 psi
<i>D</i>	diameter, feet, or top-product flow rate from condenser or condensate receiver, mols/min
<i>E</i>	Murphree tray efficiency
<i>f</i>	cycles/minute or cps
<i>F</i>	feed rate to column, mols/min
<i>g_c</i>	mass-force conversion factor, $32.2 \frac{\text{ft}}{\text{sec}^2} \frac{\text{lb mass}}{\text{lb force}}$
<i>g_L</i>	local acceleration due to gravity, ft/sec ²
<i>h</i>	heat-transfer film coefficient, $\frac{\text{pcu/sec}}{^\circ\text{C ft}^2}$
<i>H</i>	head of liquid or liquid level, feet
<i>j</i>	$\sqrt{-1}$ (has different meaning when used as subscript)
<i>K</i>	static gain
<i>l</i>	distance, feet
<i>L_o</i>	external reflux, mols/min
<i>L_R</i>	liquid downflow in column, mols/min
<i>M</i>	liquid holdup, mols
<i>MW</i>	molecular weight
<i>p</i>	pressure, psi
<i>P</i>	pressure, lbf/ft ² , or atmosphere, or mm Hg

pcu	pound centigrade units (heat required to heat one pound of water 1°C)
P_j	vapor pressure of pure component, species j
q	heat flow, pcu/sec, or fraction of feed that is liquid (molar basis)
Q	flow rate, ft ³ /sec or ft ³ /min
R	reflux ratio, L_R/D
s	Laplace transform variable
t	time, seconds or minutes
T	temperature, degrees Celsius or Kelvin, or sampling time interval in sampled-data control systems
U	overall heat-transfer coefficient, $\frac{\text{pcu/sec}}{\text{ft}^2 \text{ } ^\circ\text{C}}$
V	vapor flow, mols/min, or volume, ft ³
V_T	volume in tank corresponding to level transmitter span, ΔH_T
w	weight rate of flow, usually lbm/sec
W	weight, lbm
x	mol fraction more volatile component in a liquid
y	mol fraction more volatile component in a vapor
z	z -transform variable, or
z_F	mol fraction more volatile component in feed
Z	acoustic or hydraulic impedance, lbf sec/ft ⁵
α	relative volatility
γ	specific heat ratio, or activity coefficient
ΔH_T	liquid-level transmitter span, feet, corresponding to full-scale output
ε	difference between set-point signal and signal from measurement device
ζ	damping ratio in a quadratic expression
θ_i	arbitrary input signal
θ_o	arbitrary output signal
λ	latent heat of vaporization, pcu/lbm
λ_m	molar latent heat of vaporization, pcu/mol
μ	viscosity, lbm/ft sec = centipoise/1488
ρ	density, lbm/ft ³
τ	time constant, usually seconds or minutes
ψ	enthalpy, pcu/lbm
ω	frequency, radians/unit time

Subscripts

Q	quadratic
B	bottom of tower
R	reset, or reflux

L	light component or key
H	heavy component or key
f	feed
ff	feedforward
i	inlet
j	arbitrary tray location or component
o	outlet
s	stripping section
sp	set point
st	steam
c	controller
D	distillate (top product)
OL	open loop (used outside of brackets)

Symbols on Illustrations

CC or	composition control
XC	
FC	flow control
LC	liquid level control
PC	pressure control
TC	temperature control
HS	high signal selector
LS	low signal selector
HL	high signal limiter
LL	low signal limiter
CW	cooling water

Individual barred terms (e.g., \bar{V} , \bar{P}) indicate average values.

Combined barred terms [e.g., $\overline{HG}(z)$] have special meaning in sampled-data control systems (see Chapter 21).

$K_m G_m(s)$	measurement transfer function
$K_c G_c(s)$	controller transfer function
$K_v G_v(s)$	control valve transfer function
$K_p G_p(s)$	process transfer function

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1 Strategy for Distillation-Column Control

in chemical plants and petroleum refineries, there are, today, many distillation columns that are working well. There are also many others that are not working well, and at least a few that function very poorly, or not at all. Failure to obtain performance specified by the column design engineer is due, in many cases, to faulty or inadequate control system design. Troubleshooting of columns that are already in operation is frequently necessary, but practical considerations usually limit corrective measures to relatively minor items. Proper original design is by far the best way to guarantee satisfactory operation and control. Therefore, in this book we will approach the design of integrated distillation-column control systems as a systems problem in process design. The application of feedforward, feedback, and protective controls will be coordinated with the sizing and proper location of process holdups to achieve both automatic start-up and shutdown and smooth, noninteracting control of column product compositions.

1.1 DISTILLATION CONTROL OBJECTIVES

The starting point of any design project is a definition of objectives. For distillation there are many possible approaches, but the one chosen here is one the authors have found broadly useful in virtually all kinds of processes.¹ It has three main facets: (1) material-balance control, (2) product quality control, and (3) satisfaction of constraints. As applied specifically to distillation columns, this philosophy suggests the following:

1. Material-balance control*

* This term is sometimes used by others¹¹ to mean a control system in which reflux is set by reflux drum level control, and distillate/feed ratio is set manually or by a composition (temperature) controller. The authors of this book have been unable to find any special merit for this scheme except for some high reflux ratio columns.

—The column control system must cause the average sum of the product streams to be exactly equal to the average feed rate. Harbert²⁻⁴ has called this requirement that of keeping the column in “balance.”

—The resulting adjustments in process flows must be smooth and gradual to avoid upsetting either the column or downstream process equipment fed by the column.

—Column holdup and overhead and bottoms inventories should be maintained between maximum and minimum limits.

It is important to note that the material-balance controls on any given column must be consistent with the material-balance controls on adjacent process equipment. In most cases material balance will be controlled by so-called “averaging” liquid-level or pressure controls.

2. Product quality control

The control system for a binary distillation in most cases must:

—Maintain the concentration of one component in either the overhead or bottoms at a specified value.

—Maintain the composition at the other end of the column as close as possible to a desired composition.[†]

It is usually true that minimum operating cost is achieved when the products are controlled at minimum acceptable purities. This is so because the relationship between thermodynamic work of separation and purity is nonlinear.¹¹ For some columns compositions are allowed to vary at one end, and sometimes both ends, to satisfy certain economic constraints.

Both material-balance and composition controls must function satisfactorily in the face of possible disturbances in:

—Feed flow rate

—Feed composition

—Feed thermal condition

—Steam supply pressure

—Cooling-water supply temperature

—Cooling-water header pressure

—Ambient temperature, such as that caused by rainstorms

3. Satisfaction of constraints

For safe, satisfactory operation of the column, certain constraints must be observed. For example:

—The column shall not flood.

—Column pressure drop should be high enough to maintain effective column operation, that is, to prevent serious weeping or dumping.

[†] For multicomponent columns subjected to feed composition changes, it is not possible to hold exactly constant the compositions at both ends of the column; the composition at one end must change a little. If feed variations are in the nonkey components, composition may vary somewhat at both ends of the column. With only two drawoffs, we can control two keys, but not the nonkey components.

- The temperature difference in the reboiler should not exceed the critical temperature difference.
- Column feed rate should not be so high as to overload reboiler or condenser heat-transfer capacity.
- Boilup should not be so high that an increase will cause a decrease in product purity at the top of the column.
- Column pressure should not exceed a maximum permissible value.

There are, in addition, several other facets of column control.

Startup and Shutdown

Column controls should facilitate startup and shutdown and, by implication, should make it easy to achieve total reflux operation when desired.

Transitions

When it is desired to change product compositions, the column controls should facilitate doing so. This is particularly important when feed stock composition varies widely and it is desired to optimize column or train operation, as, for example, with a computer.

Heat Recovery

Increasingly there is an interest in recovering as much heat as possible. The petroleum industry has frequently used the sensible heat in a column bottom product to preheat column feed. Recently more ambitious schemes have been employed in which the reboiler for one column is used as the condenser for another. Such schemes magnify control problems and sometimes limit process turndown.

Testing

There should be enough instrumentation so that testing may be carried out for tray efficiency, heat and material balances, flooding, and so on.

Miscellaneous

In addition to the above, column control should be designed with human engineering in mind. For example:

- The operator's work station, whether a cathode-ray-tube/keyboard console or a panelboard with gages, switches, recorders, and so on, should be carefully designed according to human engineering principles for easy use.
- The controls should be so designed as to require minimum maintenance. The need for frequent or critical "tuning" should be avoided. The hardware should be designed and arranged for convenient access and quick repair or replacement.
- The control system design engineer should use the simplest possible design procedures, not only to hold design costs down, but also so that the work can be readily discussed with other design and plant personnel. This will facilitate, at a later date, any necessary minor redesign at the plant site. Failure

of design and plant personnel to achieve a mutual understanding of design objectives, concepts, and methods is one of the frequent causes of unsatisfactory plant operation.

1.2 ARRANGEMENTS FOR MATERIAL-BALANCE CONTROL

As shown elsewhere,¹ the size and location of tanks and the concept of overall material-balance control used can have a great influence on plant investment and process control. If the design engineer uses the concept of *control in the direction opposite to flow*, tanks may be smaller and plant fixed investment and working capital can be lower than if *control in the direction of flow* is used. The meaning of these expressions is illustrated, for simple tanks with level controllers, in Figures 1.1 and 1.2.

For the last tank in a series (final product storage), the demand flow is always the shipments to a customer. As shown by Figures 1.3 and 1.4, the choice in control strategy is between adjusting the flow into the last tank (control in direction opposite to flow) or adjusting flow into the first process step (control in direction of flow). In the first case, we can easily use simple automatic controls. In the second case, it is more common to have an operator make the adjustment.

When more than one tank is involved, other advantages of control in the direction opposite to flow are (1) less difficulty with stability problems, and (2) reduced internal turndown requirements. "Turndown," as used here, is the ratio of maximum required flow rate to minimum required flow rate. In this instance the meaning is that, in response to a given change in demand flow, the required change in the manipulated flows will be smaller in one case than in the other.

Once the basic concept of material-balance control has been selected for a process, one must apply the same concept to all process steps. It is for this reason that the first step in designing column controls is to determine the material-balance control arrangement. Control in the direction of flow is the most commonly used concept (although the least desirable), and a frequently encountered arrangement is shown on Figure 1.5. Here level in the condensate receiver (also commonly called reflux drum or accumulator) sets the top product, or distillate flow, while the level in the base of the column sets the bottom product flow; in other columns base level sets steam or other heat-transfer media to the reboiler, in which case the condensate receiver level sets top product flow.

Generally speaking the direction of material-balance control is determined by the demand stream. In recycle systems we may find some material-balance controls in the direction of flow while others are in the direction opposite to flow.²⁵

Material-balance control, in the direction opposite to flow, can lead to many interesting level-control and flow-ratio options. These are discussed in detail in Chapter 6.

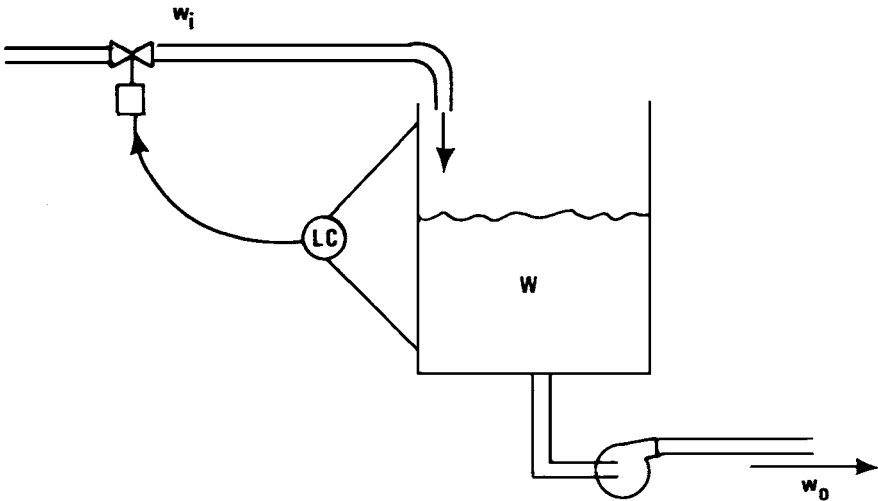


FIGURE 1.1
Material balance control in direction opposite to flow

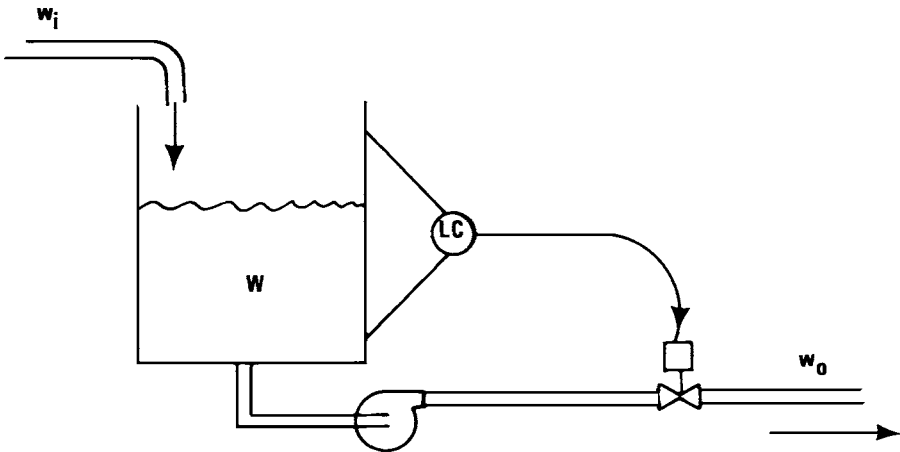


FIGURE 1.2
Material balance control in direction of flow

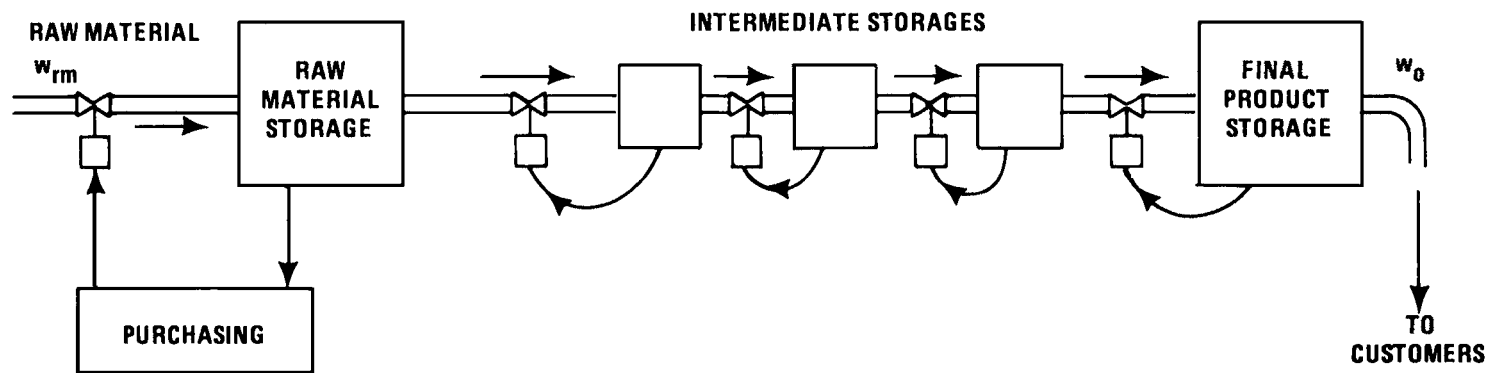


FIGURE 1.3
Overall material balance control in direction opposite to flow

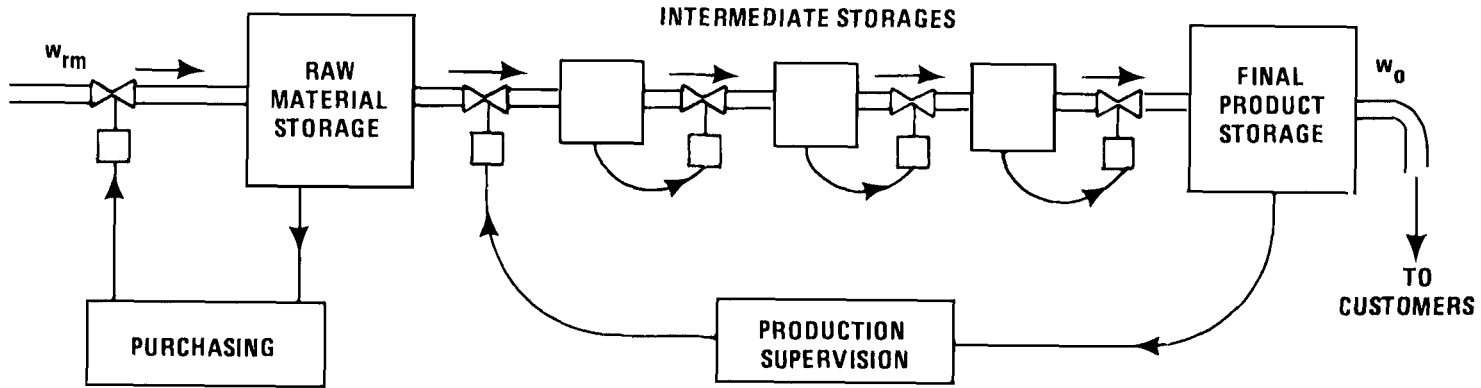


FIGURE 1.4
Overall material balance control with intermediate material balance control in direction of flow

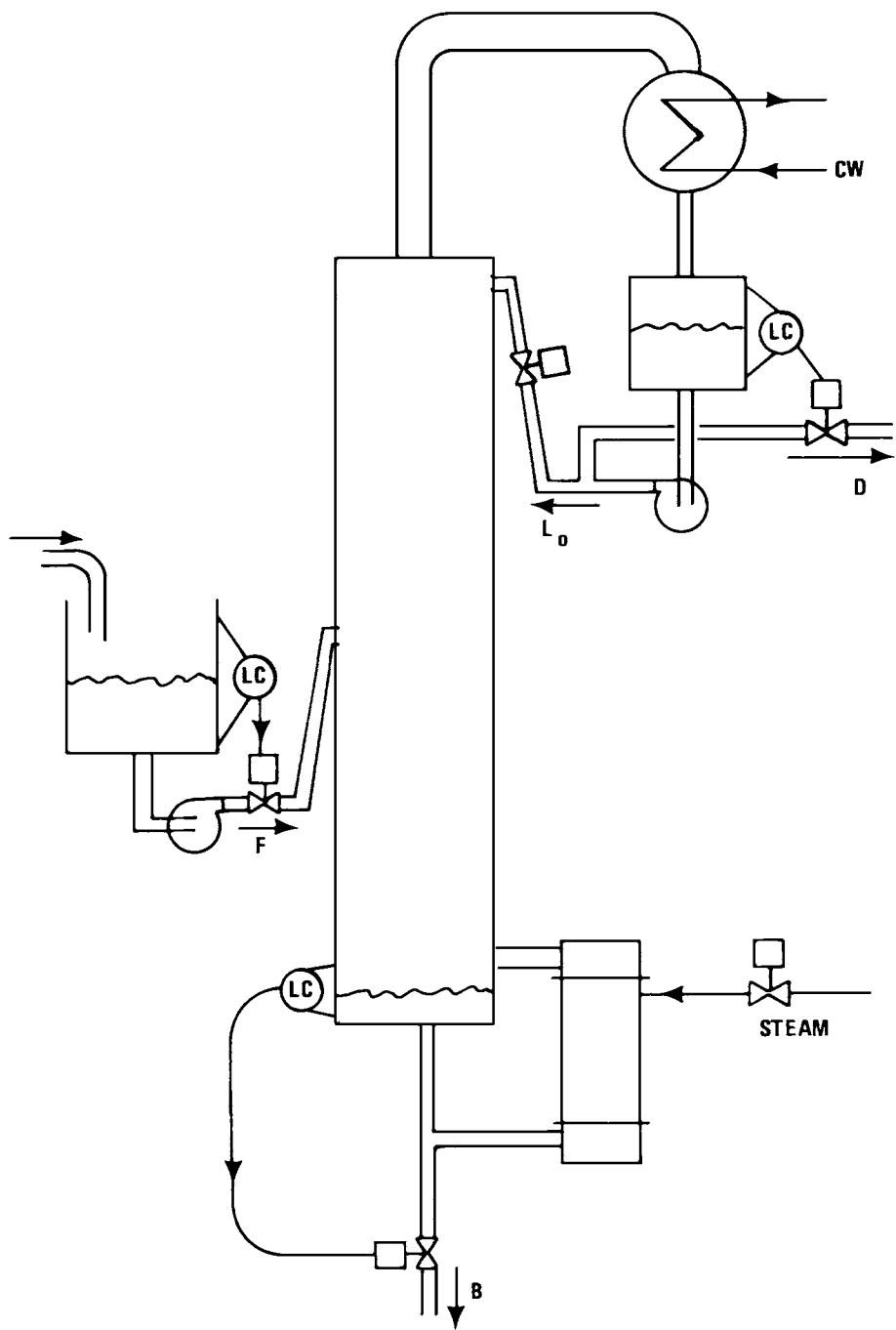


FIGURE 1.5
Distillation column with material balance control in direction of flow

1.3 FUNDAMENTALS OF COMPOSITION CONTROL

Let us consider briefly what must be done to a column to keep terminal compositions constant on a steady-state basis when the column is subjected to sustained changes in feed flow rate or feed composition. Methods of handling other disturbances will be discussed later.

To simplify the analysis, let us limit our attention to an ideal, binary distillation. This is somewhat restrictive, although the results will be applicable in a general way to multicomponent systems, particularly those that may be treated as quasi-binary or pseudobinary systems.

As will be shown in Chapter 2, if feed composition and feed thermal condition are constant, then we want the "operating lines" on a McCabe–Thiele diagram to remain constant when feed flow changes. The operating lines (defined in the next chapter) will not change as long as the distillate-to-feed, reflux-to-feed, boilup-to-feed, and bottom-product-to-feed ratios are held constant. Practically speaking one may hold all four ratios constant by fixing any one of the three pairs: (1) the reflux-to-feed ratio and the boilup-to-feed ratio, (2) the reflux-to-feed ratio and the bottom-product-to-feed ratio, and (3) the boilup-to-feed ratio and the distillate-to-feed ratio. In considering case (1), for example, we see that if the rectification section vapor-to-feed ratio is fixed (and it will be if the boilup-to-feed ratio is fixed) and the reflux-to-feed ratio is fixed, then the distillate-to-feed ratio will be fixed, since the distillate flow is the difference between the reflux flow and the vapor flow. Similarly, the bottom-product-to-feed ratio will be fixed, since the bottom product flow is the difference between the stripping section liquid and the boilup.

If feed rate and feed composition are not constant, Rippin and Lamb⁵ have shown that, for small perturbations, one should change the boilup and reflux according to the following equations:

$$\Delta V = K_{f1}\Delta z_F + K_{f2}\Delta F$$

$$\Delta L_R = K_{f3}\Delta z_F + K_{f4}\Delta F$$

where

V = vapor flow from the reboiler

L_R = internal reflux flow at the
top of the column

The Δ 's represent departures from average operating conditions. The constants $K_{f1} \cdots K_{f4}$ may be calculated approximately by the column designer. Luyben⁶ has shown that it is necessary to be quite careful in designing feedforward compensation for feed composition changes, particularly when the column is not making a sharp separation.

Visualization of column operation, in terms of reflux-to-feed and boilup-to-feed ratios, was suggested by Uitti⁷ and has since been proposed in varying degrees by many others. Throughout the rest of this book, it will be used as the primary basis for column composition control.

It should be noted, however, that this “feedforward” approach to column control has a particular limitation: In general one cannot calculate the constants $K_{f1} \cdots K_{f4}$ with great accuracy. For columns that are not operating too close to either upper or lower limits of capacity, small changes in feed rate, and consequent changes in boilup and reflux, will not change tray efficiency appreciably. The terms K_{f2} and K_{f4} therefore will be constants. If the feed composition changes are not too large (as will usually be the case), then K_{f1} and K_{f3} may also be treated as constants. To determine the control accuracy obtainable by this approach, one should make the necessary calculations or tests for each individual column. Where really close control is required, one must supplement feedforward control with measurement of the column terminal compositions and subsequent feedback control, at least at one end of the column. The usual philosophy will be to use feedforward for fast, approximate control and feedback for long-term, accurate control of composition.

It should be noted, too, that feedforward from feed composition may not be needed if the feed comes from a process step with discharge composition control. Feedforward compensation for other process variables, such as bottom product or distillate demand flow, is discussed in Chapter 6.

As will be shown later (Chapters 5 and 20), a properly designed column feed system can play a very important role in filtering out disturbances in feed rate, feed composition, and feed enthalpy, thereby making composition control much easier.

1.4 COMPENSATION FOR VARIOUS DISTURBANCES

Feed Thermal Condition

Feed should enter the column with a constant enthalpy. When significant changes are anticipated, a heat exchanger and feed-enthalpy control system should be provided. This is discussed in more detail in Chapters 5 and 11.

Steam Supply Pressure

Changes here can cause changes in boiling rate. The best preventive is a steam-to-feed ratio control system combined with temperature and pressure compensation of steam flow. A high pressure drop across the steam valve favors smooth control but velocity-limiting trim may be required to minimize noise and plug and seat wear. A high pressure drop is also undesirable for energy conservation. If the pressure drop is high enough, sonic flow through the valve results, and reboiler steam-side pressure has no effect on flow rate. The design should avoid having sonic flow corresponding to low feed rates and nonsonic flow corresponding to high feed rates, since required controller gain changes make tuning very difficult.

Cooling-Water Supply Temperature

Cooling-water temperature changes are usually seasonal, and will require no specific correction. If, for some reason, they are large and rapid, then it may be desirable to provide an enthalpy control system for the condenser. By measuring the temperature rise of the cooling water across the condenser, and multiplying it by the cooling-water flow rate, one has a measure of the heat transferred, q_c . This calculated value of q_c can serve as the measured variable in an enthalpy control system. For column pressure control, the enthalpy control system can serve as the secondary loop in a cascade system.

Cooling-Water Header Pressure

One of the best ways to make a flow system immune to pressure changes is to provide a high system pressure drop. This, however, is costly. Another way, which should be satisfactory for some distillation columns, is to use a cooling-water flow control system. It does, however, have limitations, which will be discussed later.

Ambient Temperature

If the column, auxiliaries, and piping are properly insulated, and if the column is properly controlled, ambient changes should cause little difficulty unless the condenser is of the air-cooled type. In this event it may be desirable to use an internal reflux computer, discussed in Chapter 11. If, as is often the case, the vapor piping from the top of the column to the condenser is both uninsulated and long, ambient temperature changes may cause fluctuations in pressure and the rate of condensation.

1.5 STARTUP AND SHUTDOWN

Startup and shutdown are often dismissed as relatively unimportant, since they happen so seldom that it is not economical to spend much time and money on improvements. This may be true in a petroleum refinery where shutdowns may occur at intervals of two or three years. In the chemical industry, however, where process equipment is often plagued by severe corrosion or by plugging process materials, startups and shutdowns are far more common—perhaps monthly, weekly, or even daily. Further, elaborate heat and material recycle schemes may require intricate startup/shutdown procedures as part of the original design. To put it another way, columns and their control systems may have to be designed specifically to accommodate a particular startup/shutdown procedure.

Columns are commonly started up in total reflux—no product is taken off at top or bottom. Limited experience, however, suggests that faster and smoother

startups may be achieved by recycling top and bottom products back to feed during part of the startup sequence.

Startup/shutdown will be discussed in more detail in Chapter 9.

1.6 CONTROL SYSTEM DESIGN PHILOSOPHY

Current Design Practices

In the preface we noted that we would try, in this book, to present a multivariable control approach to distillation column control. Before discussing this, however, let us look at typical controls in existing plants.

There is a traditional pattern of what is called “instrumentation” in the chemical and petroleum industries based on single-loop control (sometimes called SISO—single input, single output). Each process operation has a number of independent or single loops for feedback control of temperatures, pressures, flows, liquid levels, and sometimes compositions. The term “single loop” means there is one measurement, one controller, and one final control element, usually a valve. Many, but not all, of these loops are represented in the central control room (CCR) by control stations.

The controllers usually have neither antireset windup nor automatic tracking, and there is little or no logic circuitry to tie the many loops together. This statement is true for both analog and some digital hardware. As a consequence the operators must perform startup operations with the control stations switched to “manual,” and must implement process logic by switching in and out of “automatic.” Since the original design procedures are usually qualitative and intuitive, with heavy emphasis on conformity to past practices, it is not surprising that some loops never work in “automatic.” Others, although stable in “automatic,” are so sluggish that they are ineffective in dealing with typical disturbances. For newer plants with higher throughputs, operation closer to hard constraints, smaller and fewer holdups, energy-recovery systems, and elaborate material-recycle systems, the traditional “instrumentation” approach to control is often seriously inadequate.

For some years now, progress in single-loop design has been essentially at a standstill. As recently as 1950, process control was hardware limited. Since then primary measuring elements, controllers, computing devices, control stations, and control valves have improved greatly in reliability, sensitivity, and speed of response. Consequently these characteristics less and less frequently pose limitations to the single-loop designer. Further, the quality of control achieved by single-loop systems is not greatly affected by type of hardware; it makes little difference, for a given algorithm, whether one uses analog pneumatic or electronic gear, microprocessor controls, or a digital computer.* In addition,

* Digital computers and microprocessor controls, however, usually offer a wider range of controller parameter adjustments and facilitate the design of control systems more sophisticated than most of those discussed in this book.

optimized tuning procedures for unaided feedback controllers have limited practical value for continuous processes; they yield results that are far inferior to those obtainable with well-damped feedback controls with simple feedforward and override control.⁸

For plants being designed today (late 1983), it is increasingly common to use microprocessor controls instead of analog (see discussion under “Hardware Conventions and Considerations”).

Multivariable Control

To avoid the limitations of single-loop design and to provide a more flexible and sophisticated process operating logic than can be implemented by human operators, we use an approach we call multivariable control.⁹ Many definitions of this term may be found in the literature, but most of them are expressed in mathematical terms rather than in terms of process functions. For our purposes we define a multivariable control system as one that has the built-in intelligence to look simultaneously at two or more process variables and to choose, in a given situation, the best of several preprogrammed strategies (algorithms) for manipulating one or more control valves (or other final control elements).

For example, the steam valve for a distillation column reboiler, depending on circumstances, may respond to controllers for:

- Steam flow rate
- Column ΔP
- Column pressure
- Base temperature
- Column feed rate
- Column base level
- Column bottom-product rate

The seven variables listed may also exert control on five or six other valves.

To provide automatic control of this sort, we make extensive use of “variable configuration” controls that are usually implemented by overrides. If, for instance, base composition is normally controlled by steam flow that can be taken over or overridden by high column ΔP , this is a variable configuration. If base level control is normally achieved by a PI controller that can be overridden by high or low base level proportional-only controls, we call that “variable structure.” Multivariable control may involve both variable configuration and variable structure controls. Hardware permits us to automate this kind of control with a speed, precision, and reliability that are completely beyond the capabilities of human operators.

It is common to think of process control functions stacked one above another in a pyramid or hierarchic arrangement as with traditional business or military organization structures, particularly when computers are involved. Multivariable control structures, however, with their extensive lateral and diagonal crossovers, are functionally more like the modern “matrix” concept of management. From

a control engineering standpoint, the shorter lines of communication and decentralized control functions permit more rapid and stable control, and more reliable, troublefree operation.

By now it is probably apparent that we are striving for control system designs whose performance and design parameters are specified in advance of plant startup. In practice we furnish calibration data for controller parameters and computational devices for the majority of control loops prior to startup. We calculate these from simulations or simple linear models. For microprocessor computer controls, we calculate scaling parameters for computation blocks (either in software or hardware). Our design procedures are accurate enough that only a modest amount of empirical controller tuning is required at startup.

Stability, Speed of Response, and Interactions

Most existing literature on automatic control is concerned with the stability and speed of response of single loops. The traditional objectives of feedback control system design are:

1. To get the fastest possible response to set point changes.
2. To compensate for or to attenuate disturbances as much and as quickly as possible. These must be accomplished with a reasonable degree of closed-loop stability.*

In process control the objectives are often quite different. The objectives of averaging level control, for example, clearly are different from those just mentioned. A typical chemical plant or refinery has hundreds of single loops with many interactions among them. It is usually far more important to design for a dynamic balance among these loops with a minimum of interaction than to strive for maximum speed of response. Further, it is usually undesirable to make rapid changes in manipulated variables since these may upset the process. For example, distillation column reflux flow and boilup should not be changed too rapidly since these might cause transient flooding or weeping in part of the column. Our preferred philosophy of controller tuning is discussed in a book¹ and a paper.¹⁰

There are five simple methods by which to make a system noninteracting:

1. Design material-balance control loops to be at least a factor of 10 slower than related composition control loops. Similarly, in cascade systems, make the secondary or slave loop at least a factor of 10 faster than the master loop.
2. Avoid designs that are intrinsically interacting, such as pressure control and flow control at the same point in a pipeline. One of the two controllers must be detuned.
3. Select process designs that eliminate or minimize interaction. For example,

* Some processes are open-loop unstable, which means that controls must be added and must be kept in "automatic" for stable operation.

if a tubular reactor is fed at several points along its length from a common header, flow-rate interactions may be reduced to an arbitrary level by providing a very high pressure drop across each feed valve in comparison with the reactor pressure drop.

4. Use override circuits. Although not specifically intended for this purpose, override circuits, by permitting only one controller at a time to control a given valve, eliminate interactions.

5. Use interaction compensators (decouplers). If two control loops, such as top and bottom temperature controls on a distillation column, interact, we can eliminate the interactions by installing two compensators. One compensates for the action of the top temperature controller on the bottom loop while the other compensates for the action of the bottom temperature control loop on the top one. This is discussed in Chapter 20.

In addition, there are some very sophisticated mathematical methods for dealing with interactions.^{23,26,27} Some are intended for noninteracting design while others seek a design that provides an optimum amount of interaction.

Hardware Conventions and Considerations*

For control loops represented in the CCR (central control room), it is normal practice (as mentioned earlier) to furnish “control stations.” These may be analog pneumatic, analog electronic, or microprocessor based. In the last case, the station may be physically distinct, like an analog station, or may be represented on a CRT display as a “faceplate.” Each provides an indication of the process variable (flows, level, temperature, etc.), the desired value (set point), and the valve loading signal (controller output, really). There is also a “manual-automatic” switch, which some vendors label “hand-automatic.” In the “manual” mode, the feedback controller is disconnected and there is a knob that enables the operator remotely to set the valve position. This may or may not be subject to restrictions imposed by feedforward compensation, overrides, and so on, depending on the design philosophy for a particular project.

If cascade control is involved, as, for example, liquid level control cascaded to flow control, the secondary station not only has manual-automatic switching, but also another function—“remote-local.” In the “remote” position, the secondary controller set point comes from the output of the primary controller. In the “local” position, valve position may be set manually or the controller set point may be set by the operator (“local-auto”). Although cascade functions are sometimes combined into one station (or CRT “faceplate”) for space and money-saving reasons, we recommend dual stations. Most single-station designs with which we are familiar are very inflexible and complicated; they do not permit ready implementation of feedforward, overrides, and so forth.

* Much of this section may seem superfluous to instrument and plant personnel. Our experience, however, with process design engineers, column designers, and so forth is that this discussion may be very helpful.

As of this writing (late 1983), the two biggest hardware needs appear to be:

1. Better measurements, especially of compositions.
2. Better control valves. With regard to these, more progress has been made in the design of valve bodies and trim than in the design of actuators. Valve positioners should always be used (it is assumed in this book that they will be).^{*} Piston actuators with double-acting, two-stage positioners are recommended.

As far as CCR hardware is concerned, we have a decided preference for microprocessor controls. They are technically more versatile and are less expensive (some versions) than analog. As of late 1983, many are featuring satisfactory antireset windup and override capabilities. In addition, they provide more advanced logic capability, dead-time simulation, and adaptive tuning. Some of the last named achieve self-tuning via stochastic techniques or by pattern recognition. Others have gain scheduling, where reset time and proportional gain are functions of some process variable or the controller error signal.

Microprocessor controls usually have a sampling time of a fraction of a second. Although slightly slower than analog controls, their performance can generally be approximated by analog control algorithms.

Other advantages include freedom from drift, and the fact that they can be calibrated more precisely, can be reconfigured or restructured without wiring changes, have a larger range of tuning parameters, and contain more control algorithms.

For most projects today, it is possible to find worthwhile applications for a supervisory digital computer with a good data historian, regardless of the type of basic controls selected (pneumatic, electronic, or microprocessor). Digital readouts for important variables are worthwhile because they permit seeing their magnitudes with sensitivity approaching that of the original analog measurements. Most typical analog measurements have a sensitivity ranging from one part per 1000 to one part per 10,000. Most analog readout devices, however, are limited to 0.5–2 percent.

For maximum advantage a supervisory computer should be programmed to have the control algorithms discussed in Chapter 12. These are position rather than velocity algorithms. It is our opinion that using such a computer to imitate unenhanced two- and three-mode analog controls is poor practice. Some worthwhile applications for computers will be discussed later.

Computer consoles were originally provided in the CCR for the convenience of the operators. Sometimes a consolidated console is also provided for production supervision. Engineers' consoles, perhaps at another location, facilitate technical studies. Separate consoles for maintenance personnel (the computer can be a powerful maintenance tool) are highly desirable.

^{*} Most control valves today are either single-seated global types or rotary types. For both there are substantial, coercive stem forces when the valves are in flowing streams. Positioners compensate for this and maintain the valve's inherent flow characteristic expressed as a function of controller output signal. It should be noted, however, that some users prefer not to use positioners.

1.7 PROCEDURE FOR OVERALL CONTROL SYSTEM DESIGN

We now have enough information to suggest a sequence of design steps to follow that will lead to a quantitative definition of column and hardware performance.

1. After careful discussion with the process engineer and column designer, and after careful review of the overall process flow sheet, prepare a simplified flow sheet that defines the control concepts:

a. Select the overall material-balance control scheme first, preferably proceeding from final product inventory to raw material inventory. All individual equipment-piece material-balance controls must be consistent with this scheme.

b. Select composition control schemes.

c. Add feedforward and interaction compensators as required.

d. Add protective controls and automatic startup/shutdown circuits.

e. Add miscellaneous temperature and pressure controls.

2. Prepare material-balance and composition-control signal flow diagrams.

3. Determine holdup volumes required for smooth material-balance control and for liquid-level override controls at each end of the column.

4. With holdups determined calculate column-composition transfer functions.

5. Select measurement spans and calculate control valve sizes.

6. Calculate feedforward and interaction compensators.

7. Calculate all other overrides.

8. Calculate feedback-controller gain and reset settings and control-loop natural frequencies. Check feed-tank material balance and mixing time constants for adequacy.

9. Use simulation for some columns, particularly those in critical service or with a new untested control system or process configuration. Simulation of the column and its control system will be useful in confirming control concepts and controller tuning parameters. It may also save startup time.

Hardware vendors may now be selected and the measurement and control equipment may be ordered.

Part II of this book deals with the qualitative and heuristic aspects of steps 1 and 3. Quantitative information for the other steps is presented in Part III.

1.8 COLUMN DESIGN PHILOSOPHY AND CONTROL SYSTEM DESIGN

Experience on many projects shows that even small, simple columns benefit from a modest application of feedforward compensation and overrides. This is due to the trend toward increasingly tight column design. The number of trays is held down because smaller allowances are made for uncertainty in tray efficiency. Column diameter and tray spacing are now kept to smaller values, with the

result that columns typically operate closer to flooding. The combined effect of these design policies is to make columns much touchier and harder to control.

Experience indicates that typical incremental instrument investment over that required for unenhanced feedback controls is 5–10 percent (large projects tend toward the lower figure). This incremental investment not only provides better normal control, but it also helps to avoid inadvertent shutdowns. It is, therefore, our opinion that these controls should be used to some extent on almost every column.

Suppose, however, that the customer insists on minimum application of feedback controls with no feedforward compensation or overrides. What column design philosophy should be followed? Having had considerable adverse experience with columns with primitive controls, particularly sidestream drawoff columns, and columns with heat recovery schemes, we suggest the following:

1. Design for normal operation at 60 percent of the rate for flooding.*
2. Provide five extra trays or increase the number of trays by 10 percent, whichever is larger.
3. Provide increased tray spacing.
4. Provide larger condensers. If water cooled, use tempered water.
5. Provide larger reboilers (lower heat flux).
6. Control column ΔP by boilup, or flow control steam.
7. Provide 100 percent reserve capacity in heat-recovery schemes or avoid them altogether.
8. Avoid sidestream drawoff designs.
9. Provide surge tanks between columns with at least 30–60 minutes of holdup each.

1.9 EXISTING COLUMNS—TYPICAL PRACTICES AND TROUBLESHOOTING

Although this book deals primarily with distillation control in design projects, it is pertinent to consider briefly the controls of typical, existing columns, the opportunities for their improvement, and how to troubleshoot them when necessary. Frequently encountered problems include unstable or ineffective controls, off-specification product or products, and flooding or dumping. In addition, it is fairly common practice to use excessive boilup and reflux to make sure of meeting or exceeding product specifications. This not only wastes energy; it also reduces column capacity. To provide a perspective on energy savings, one may note that 100 lbm/hr of steam is worth \$3200 per year (basis \$4.00/1000 pounds, 8000 hours per year). To save this amount of steam would probably be only a modest accomplishment for most columns.

If composition control of each product stream is desired (and this is usually

* In some companies columns are designed by the probabilistic methods recommended by Fractionation Research, Incorporated.

the case), the most obvious deficiency of most existing columns is the lack of appropriate composition measurements. Most commonly temperature in the upper or lower section of the column (or both) is used in lieu of true composition measurements. Frequently composition control is attempted at only one end of the column, and sometimes at neither end.

Another shortcoming frequently observed is the use of fixed flow controls for steam, reflux, or product drawoff. Any such unaided flow control should be regarded with suspicion. With rare exceptions flow-control set points must be changed to accomplish either composition control or material balance control.

In view of the preceding comments about problem areas and likely opportunities for improvement of composition control and reduction in energy consumption, the following guidelines are suggested:

1. Make sure that column material-balance controls are properly designed and tuned, and that hardware, especially level and flow transmitters and control valves, is in good working condition. If PI level controllers are used, follow the tuning procedures of Chapter 16; auto overrides or nonlinear controllers should be used. It is usually desirable to cascade level control to flow control, in which case flow measurement should be linear.

2. Provide averaging level control of column feed. Column feed rate should also be held between maximum and minimum limits.

3. Provide a linear flow measurement for steam flow control. Also provide temperature and pressure compensation.

4. If condenser controls are a problem, review the control schemes in Chapter 3.

5. At this point with flows established, smoothed, and in some cases limited, it will probably be possible to see some improvement in composition control, at least for part of the time. For further improvement provide steam-to-feed ratio control, internal reflux-to-feed or distillate-to-feed ratio control, high ΔP override on steam to protect against flooding, and a minimum steam flow limiter to protect against dumping.

6. If composition is measured and is cascade controlled via reflux or boilup, or both, ratio controls should be replaced by impulse feedforward compensation (see Chapter 12) if feed flow turndown is greater than 2:1.

7. For pressurized or vacuum columns, make sure an adequate scheme is provided for maintaining an inert balance.

8. If temperature by itself is not an adequate measure of composition, consider one of the schemes in Chapter 10 for using temperature, pressure, and, in some cases, flow measurements to deduce compositions.

1.10 CONVENTIONS FOLLOWED IN THIS BOOK

Throughout the remainder of this book, we will frequently illustrate control schemes with pneumatic components. This is done for convenience. Pneumatics,

with few exceptions, features a standard signal span of 3–15 psig. All pneumatic controllers, as far as we know, can achieve antireset windup by the same external reset feedback method. This appears to be the most universally useful method (we use it extensively), and has been adopted by one manufacturer of analog electronic instruments and by several vendors of microprocessor-based distributed controls. Other vendors of electronic analog and digital controls feature a wide variety of techniques; some of these work fairly well, while some are quite inflexible. A brief discussion is presented in Chapter 12.

Units used in this book are those commonly employed in chemical engineering: pounds, feet, degrees Celsius, mols, and so on. To facilitate the calculation of control engineers' "time constants," we have mostly used time units of minutes or seconds [e.g., lbm/sec, lbm/min, (pcu/sec)/°C ft²]. For projects that make partial use of metric or SI units, we found it convenient to convert them to the above units. Generally speaking, we have found no advantage in writing equations with SI units. Instead, in programs for computers or programmable calculators, we mostly write the equations in the older units and add subroutines for going back and forth to metric or SI units.

The most common type of controller used in the chemical and petroleum industries was once called "proportional plus automatic reset," later shortened to "proportional reset." Today it is more common to use "PI," which stands for proportional-integral. It is also becoming common today to speak of controller gain rather than proportional band ($PB = 100/K_c$). We will also use "reset time," usually in minutes, rather than its older reciprocal, "repeats per minute."

For drawings in which we are trying to present a perspective of control concepts (configurations), we use very simplified symbols. In drawings where we are trying to illustrate concepts of structure (overrides, feedforward, etc.), we use a more detailed symbolism that we have found useful in our design work.

In pneumatics it is common to refer to most signal-conditioning devices other than controllers as "relays." Included are adders, subtractors, multipliers, and so on. For electronic analog and digital controls, it is more common to use terms such as *signal scalers* and *multipliers*.

A table of nomenclature and symbols will be found at the end of this book.

1.11 LITERATURE

For anyone seriously interested in distillation control, two books are highly recommended. The first is an easy-to-read, nontheoretic (as far as control is concerned) work by F. G. Shinskey.¹¹ The treatment of energy conservation alone is worth the price of the book. The second book is by Rademaker, Rijnsdorp, and Maarleveld.¹² It relies heavily on conventional, single-loop control theory, and explores painstakingly a large number of possible control systems. It also contains an extensive bibliography.

For basic reference books on distillation, we have made much use of those by Van Winkle¹³ and King.¹⁴ Others by Holland,^{15,15a} Treybal,¹⁶ and Hengstebeck¹⁷ also have been useful.

For basic books on control, we recommend two written by the authors,^{1,18} one by Harriott,¹⁹ and one by Murrill.²⁸ For more advanced treatments, we suggest texts by Koppel,²⁰ Douglas,²¹ and Gould.²² The last contains some perceptive comments about the difficulties of applying advanced control theory developed by electrical and mechanical engineers to chemical processes. Recent books by Ray²³ and by Stephanopolous²⁷ discuss applications of "modern control theory" in chemical processes. McAvoy²⁶ has addressed the specific subject of interaction analysis.

Nonchemical engineers with no background in distillation may find an introductory text by Nisenfeld and Seemann useful.²⁴ It is clearly written and easy to read.

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2

Fundamentals of Distillation

2.1 INTRODUCTION

the purpose of this chapter is to review briefly the essential features of the distillation process. The control engineer must have a basic understanding of any process before an effective control system can be developed. Most readers with chemical engineering backgrounds will be familiar with this material and can skip some sections of this chapter.

Fundamental Objectives

Distillation columns are very widely used in the chemical and petroleum industries to separate chemical components into more or less pure product streams. This separation is based on differences in “volatilities” (tendencies to vaporize) among various chemical components. For example, a mixture of methanol and water can be separated by distillation because methanol is more volatile or boils at a lower temperature than water. In a distillation column, the more volatile, or lighter, components are removed from the top of the column, and the less volatile, or heavier, components are removed from the lower part of the column.

Nomenclature

Figure 2.1 summarizes the nomenclature conventions that will be used throughout this book. At this point we will consider only a simple single-feed, two-product column separating a binary (two-component) mixture.

Feed rate is F mols per minute. Feed composition is z_F mol fraction of the more volatile component. The column trays are numbered from the base upward, with feed introduced on the N_F tray. The total number of trays in the column is N_T .

Products removed from the top and bottom of the column are called “distillate” or “top product,” and “bottoms” or “bottom product,” respectively, with flow

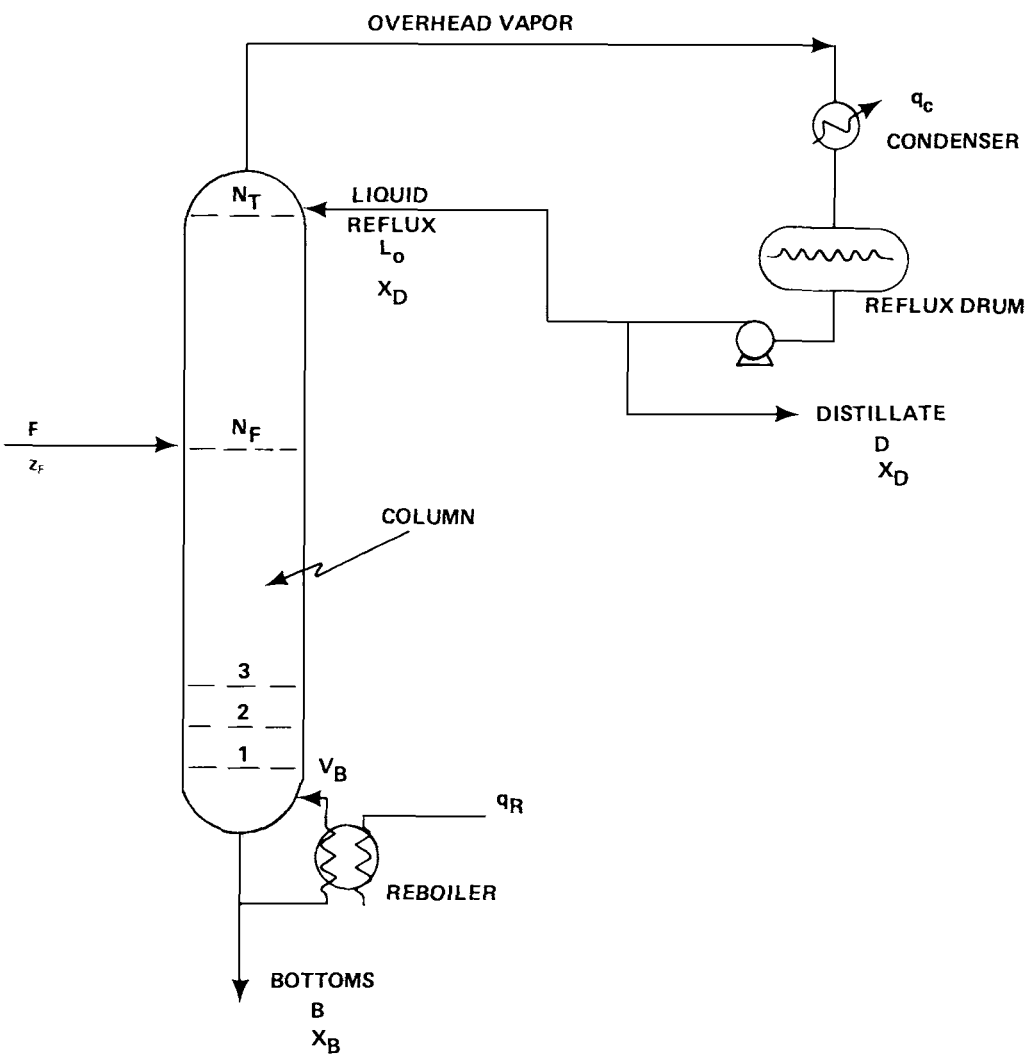


FIGURE 2.1
Nomenclature and conventions for typical distillation column

rates D and B mols/min and compositions x_D and x_B mol fraction light component.

Heat is transferred into the process in the “reboiler” (typically a tube-and-shell heat exchanger) to vaporize some of the liquid from the base of the column. The heat-transfer rate is q_R energy units/time (e.g., Btu/hr, pcu/hr, joules/minute, etc.).

The vapor coming from the top of the column is liquified in another tube-and-shell heat exchanger called a condenser. Heat is transferred out of the condenser at a rate q_c , pcu/hr.

Liquid from the condenser drops into the reflux drum. Distillate product is removed from this drum. In addition, some liquid, called “reflux” (L_o , mols/min), is fed back to the top tray of the column. This liquid reflux and the vapor boilup in the base of the column are necessary to achieve the separation or “fractionation” of chemical components. The energy required to make the separation is approximately the heat added to the reboiler.

Overall View from a Control Perspective

One can stand back and look at a distillation column with its associated reboiler, condenser, and reflux drum as a “black box” process. Feed, heat, and reflux are inputs into this box (see Figure 2.2). Outputs from the box are the two product streams D and B with compositions x_D and x_B . The usual situation with a distillation column is that the feed rate and feed composition must be

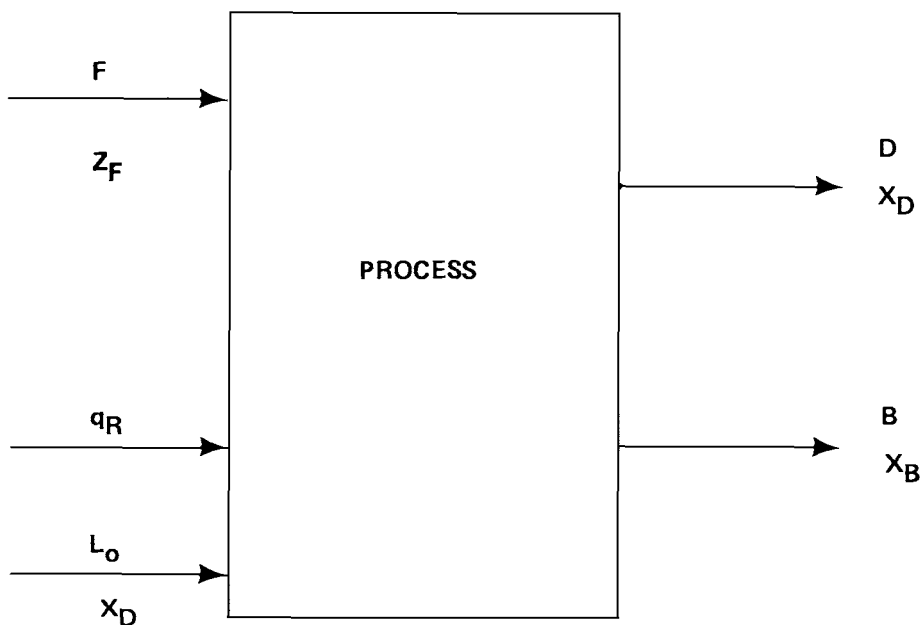


FIGURE 2.2
Control variables for distillation column

considered as disturbances. Heat input q_R and external reflux L_o can be adjusted to achieve the desired control objectives. These are called manipulated variables. Distillate and bottom product rates can also be manipulated, so there are four variables that can be adjusted: L_o , q_R , D , and B . They are not independent, however. If, for example, D is controlled, then B is dependent.

Fundamental Manipulated Variables

Two of the four manipulated variables listed above must be used to maintain liquid inventories in the reflux drum and in the column base. Therefore, we are left with just *two* manipulated variables that can be used to control compositions in the column.

No matter what “manipulated variables” are chosen to control what “controlled variables,” there are basically two fundamental manipulated variables that affect compositions. These are “feed split” and “fractionation.”

Feed split means the fraction of the feed removed as either distillate or bottom product. The D/F and B/F ratios can be manipulated either directly (as proposed by Shinskey in his “material balance control” scheme) or indirectly. The steady-state effectiveness of both the direct and indirect schemes is identical. In either case feed split has a very strong effect on product composition. A slight change in feed split can change product compositions very drastically, particularly when product purities are high.

Fractionation also affects product composition. Fractionation means the degree of separation. It varies with the number of trays in the column, the energy input to the reboiler, and the intrinsic difficulty of separating the components. For a fixed column operating at a fixed pressure with given chemical components, heat input to the reboiler is the only variable that can be used. Heat input can be used directly; alternatively reflux can be adjusted, if this is more convenient, since reflux and heat input are tied together through overall energy and mass balances.

We will discuss the pros and cons of various choices of control schemes in more detail in later chapters.

2.2 TRAY HYDRAULICS

The vast majority of industrial distillation columns are equipped with trays or plates (sometimes called “decks” in the petroleum industry) located every 1–3 feet up the column. These trays promote mass transfer of light components into the vapor flowing up the column and of heavy components into the liquid flowing down the column. Vapor–liquid contacting is achieved by a variety of devices. The most widely used trays in recent years have been sieve trays and valve trays because of their simplicity and low cost.

Sieve trays are simple flat plates with a large number of small holes. Vapor flows up through the holes, preventing the liquid from falling through. Liquid

flows across each tray, passes over a weir, and drops into a “downcomer,” which provides liquid for the tray below through an opening at the base of the downcomer. See Figure 2.3. Valve trays are built with a cap that fits over the hole in the tray and that can move up and down, providing more or less effective hole areas as vapor flow rate changes.

This fairly complex process of flow of vapor up the column and of liquid down the column is called tray “hydraulics.” It is important in control system design because it imposes very important constraints on the range of permissible liquid and vapor flow rates. If liquid cannot flow down the column, or if vapor–liquid contacting is poor, the separating ability of the column drops drastically.

Vapor flows from one tray up through the tray above it because the pressure is lower on the upper tray. Thus there is an *increase* in pressure from the top of the column to its base. Liquid must flow against this positive pressure gradient. It is able to do so because the liquid phase is denser than the vapor phase. A liquid level is built up in the downcomer to a height sufficient to overcome the difference in static pressure between the tray onto which the liquid is flowing and the tray from which it is coming.

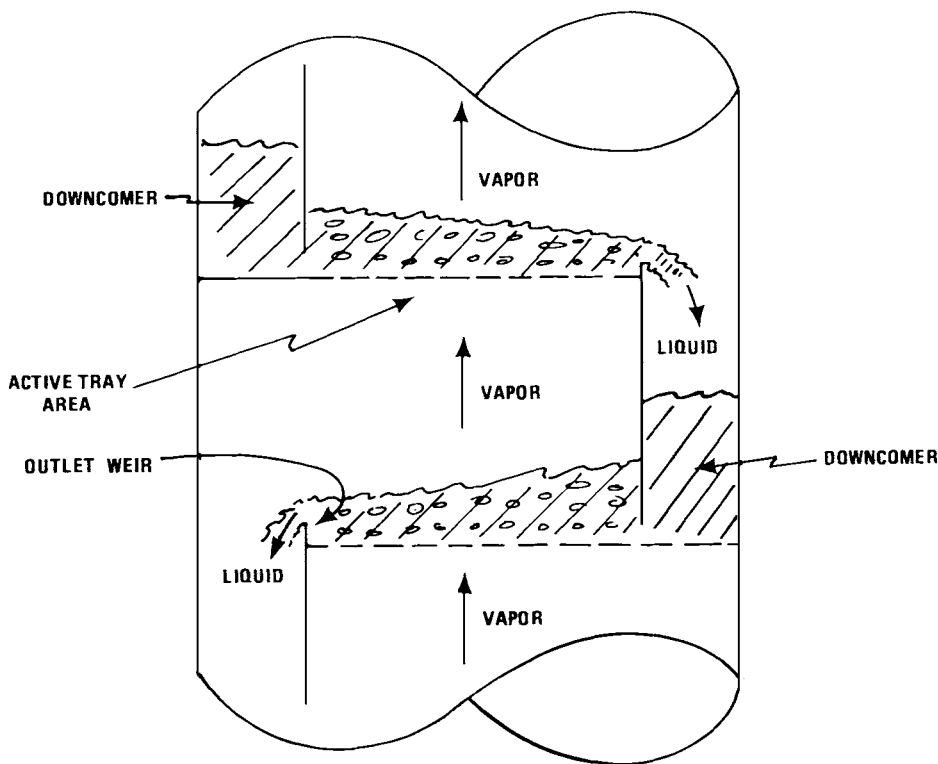


FIGURE 2.3
Schematic of typical sieve tray

This pressure difference depends on the vapor pressure drop through the tray (which varies with vapor velocity, number and size of holes, vapor density, etc.) and the average liquid height on the tray (which varies with liquid flow rate, outlet weir height, etc.).

Tray “flooding”^{*} occurs when the liquid height in the downcomer equals or exceeds the height between trays (tray spacing). This is usually due to excessive boilup (vapor rate) but sometimes may be caused by excessive reflux. The control system must keep the column from flooding. Therefore, there are maximum vapor and liquid rates.

On the other end of the scale, if vapor rates are reduced too much, the vapor pressure drop through the openings in the tray will be too small to keep the liquid from weeping or dumping down through the holes.[†] If this occurs, vapor–liquid contacting is poor and fractionation suffers. The same thing occurs if liquid rates are so low (as they often are in vacuum columns) that it becomes difficult to hold enough liquid on the tray to get good vapor–liquid contacting.

These hydraulic constraints can be handled in control system design by using maximum and minimum flow limiters on heat input and reflux. A measurement of column pressure drop can also be used to prevent flooding.

2.3 VAPOR–LIQUID EQUILIBRIUM FUNDAMENTALS

Distillation columns can be used to separate chemical components when there are differences in the concentrations of these components in the liquid and vapor phases. These concentration differences are analyzed and quantified using basic thermodynamic principles covering phase equilibrium. Vapor–liquid equilibrium (VLE) data and analysis are vital components of distillation design and operation.

Vapor Pressure

The liquid phase of any *pure* chemical component, species j , exerts a certain pressure at a given temperature. This pressure is called the pure component “vapor pressure” P_j . It is a physical property of each component.

Vapor-pressure data are obtained by laboratory experiments where both liquid and vapor phases of a pure component are held in a container (see Figure 2.4). Pressure is measured at various temperatures. The temperature at which the pure component exerts a pressure of one atmosphere is called its “normal boiling point.” Light components have low normal boiling points and heavy components have high normal boiling points.

^{*} For a further discussion of flooding, see pages 424–430 of reference 8.

[†] This occurs at about 60 percent of design vapor rates for sieve trays and about 25 percent of design vapor rate for valve trays.

When the data are plotted on linear coordinates (see Figure 2.5), a nonlinear dependence of vapor pressure on temperature is obtained. Vapor-pressure data often can be described by the Antoine equation*:

$$\ln P_j = A_j + B_j/T \quad (2.1)$$

where

P_j = vapor pressure of j th component in any pressure units
(commonly mm Hg, psia, atmospheres, kPa)

T = absolute temperature (degrees Kelvin or Rankine)

A_j and B_j = constants over reasonable range of temperatures

Therefore, vapor-pressure data are usually plotted using coordinates of log pressure versus reciprocal of absolute temperature as illustrated in Figure 2.6.

Note that the constants A_j and B_j must be determined for each pure component.

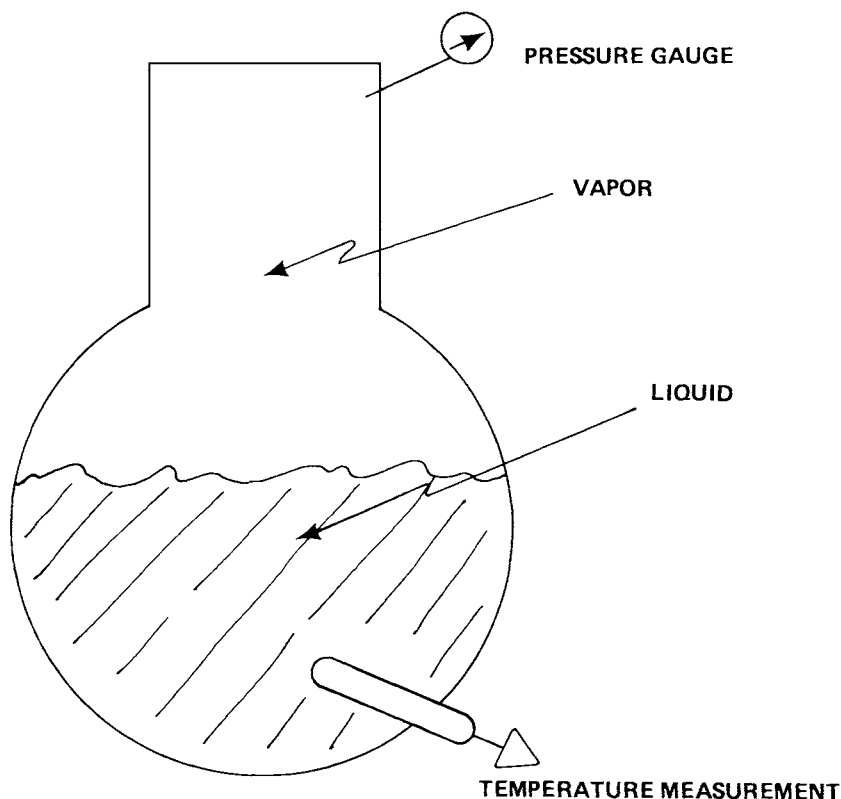


FIGURE 2.4
Vapor pressure and temperature measurement

* A three-constant version of the Antoine equation is used in Chapter 10.

They can be easily calculated by knowing two vapor-pressure points (P_1 at T_1 and P_2 at T_2). When working with a distillation column, T_1 and T_2 are usually selected to be near the temperatures at the top and at the bottom of the column.

$$B = \frac{(T_1)(T_2)}{T_1 - T_2} \ln \frac{P_2}{P_1} \quad (2.2)$$

$$A = \ln P_2 - \frac{B}{T_2} \quad (2.3)$$

Experimental VLE Data for Binary Systems

The VLE data for binary systems are obtained experimentally by mixing two components and allowing the vapor–liquid system to equilibrate. Then temperature and pressure are measured. Samples of the vapor phase and liquid phase are taken and analyzed.

Liquid compositions are usually expressed as mol fraction of light component and the symbol x is used. Vapor composition is expressed as mol fraction light component, using the symbol y . Mixture composition is then changed and the procedure repeated.

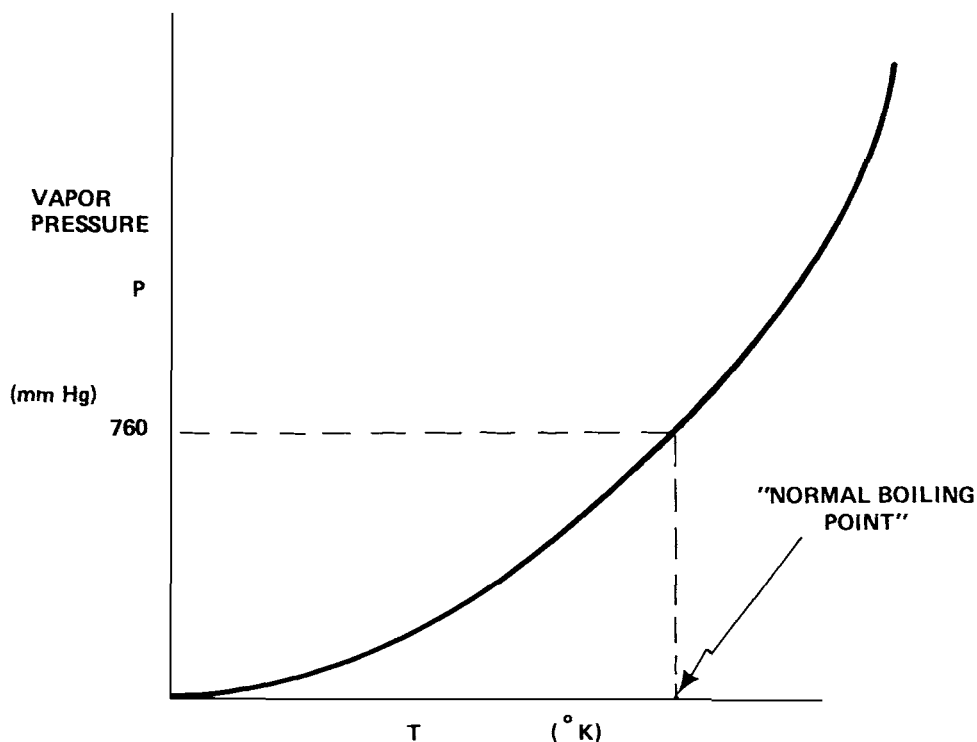


FIGURE 2.5
Temperature vs. pressure for pure component

These results are conveniently presented in graphic form using several types of "phase diagrams."

A. T-xy Diagrams

If the data are taken with pressure held constant (isobaric), it is convenient to plot two curves on the same paper: temperature versus x and temperature versus y . Figure 2.7 illustrates a typical T-xy diagram. To determine the composition of liquid and vapor phases in equilibrium with each other at a given temperature, and at the pressure under which the data were obtained, one merely draws a horizontal line at the given temperature and reads off x and y values.

B. P-xy Diagrams

Data taken at constant temperature (isothermal) are plotted as two curves: pressure versus x and pressure versus y , as illustrated in Figure 2.8. Note that the ends of the curves must intersect at $x = y = 0$ and at $x = y = 1$, since these points correspond to pure components, either light or heavy.

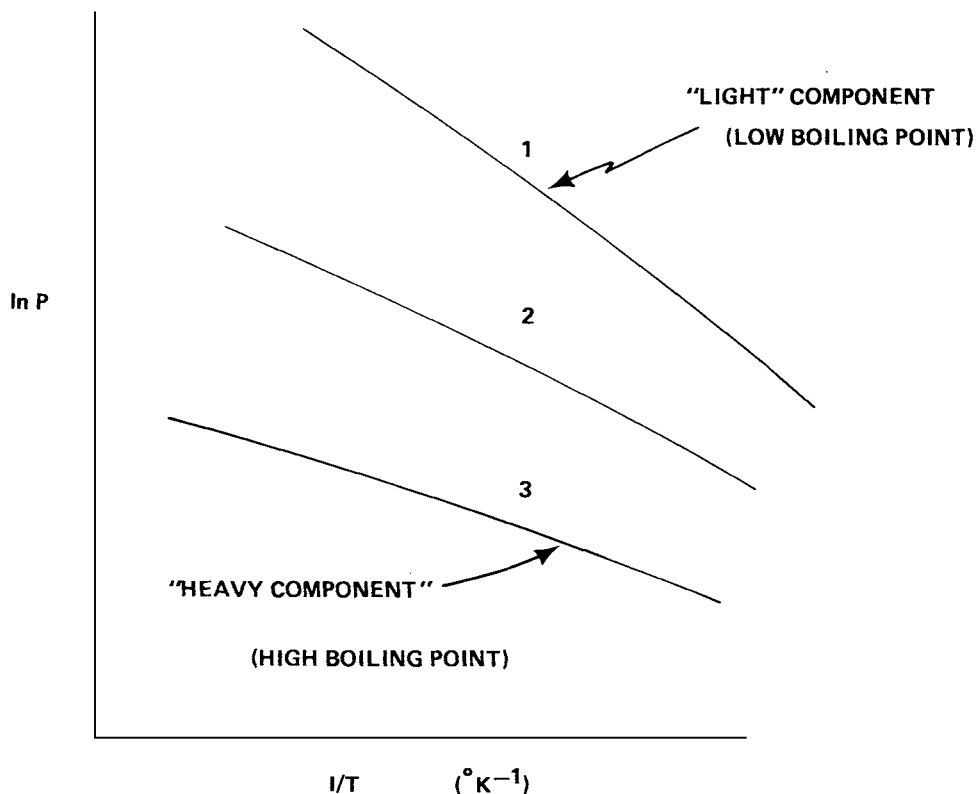


FIGURE 2.6
Typical method of plotting vapor pressure vs. temperature

C. x - y Diagrams

Either isothermal or isobaric data can be represented by simply plotting liquid composition x versus vapor composition y . This x versus y curve (see Figure 2.9) is called the “equilibrium line.” This type of diagram is the most widely used on distillation.

Both T - xy and P - xy are useful in illustrating the concepts of bubble point, dew point, superheated vapor, and subcooled liquid. Consider the P - xy diagrams sketched in Figure 2.10. Suppose we have a mixture with composition z , and we hold it at the same temperature T for which the diagram was drawn. If we impose a very high pressure on the mixture, we will be above the x versus P curve (saturated-liquid line) and no vapor will be present. There will be only

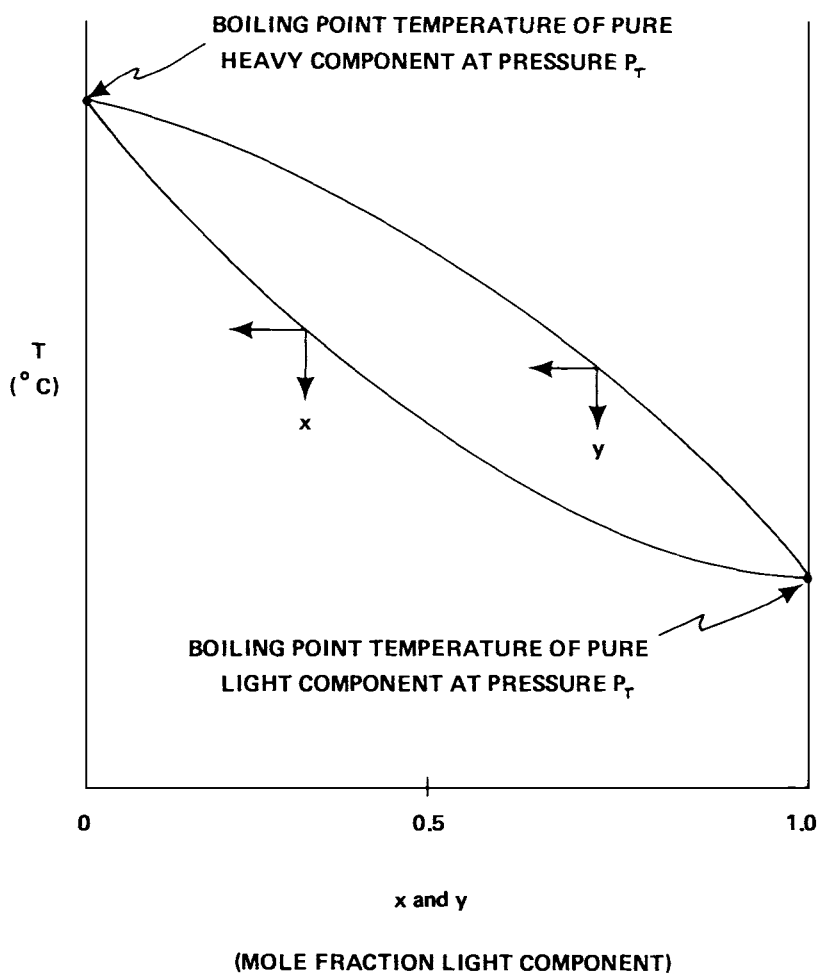


FIGURE 2.7

Temperature vs. composition of binary mixture at constant pressure

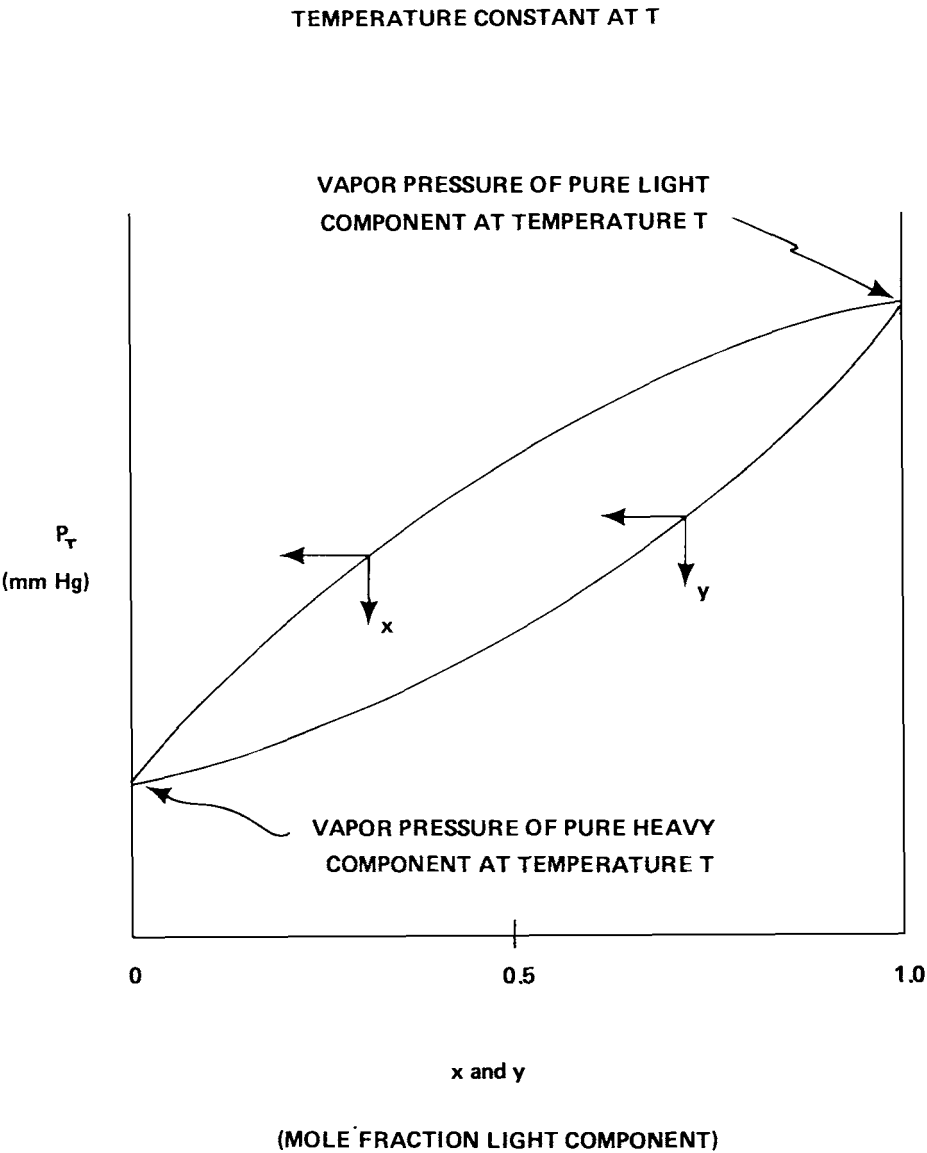


FIGURE 2.8
Pressure vs. composition of binary mixture at constant temperature

a liquid phase present with composition z . At this high pressure, the liquid is called "subcooled."

If we now begin to drop the pressure, we will move down the vertical line drawn through composition z , shown in Figure 2.10. When pressure reaches the point labeled P_{BP} , vapor will begin to appear. This therefore is called the bubble-point pressure of this mixture of composition z and at temperature T . The composition of this first bubble can be read off the y versus P curve by moving across horizontally at P_{BP} .

As pressure is reduced further, more and more vapor is formed. Finally, at a pressure P_{DP} all the liquid has vaporized. This is called the dew-point pressure of this mixture of composition z and at temperature T . The y versus P line is called the saturated-vapor line. At pressures below P_{DP} only a single phase exists, "superheated vapor."

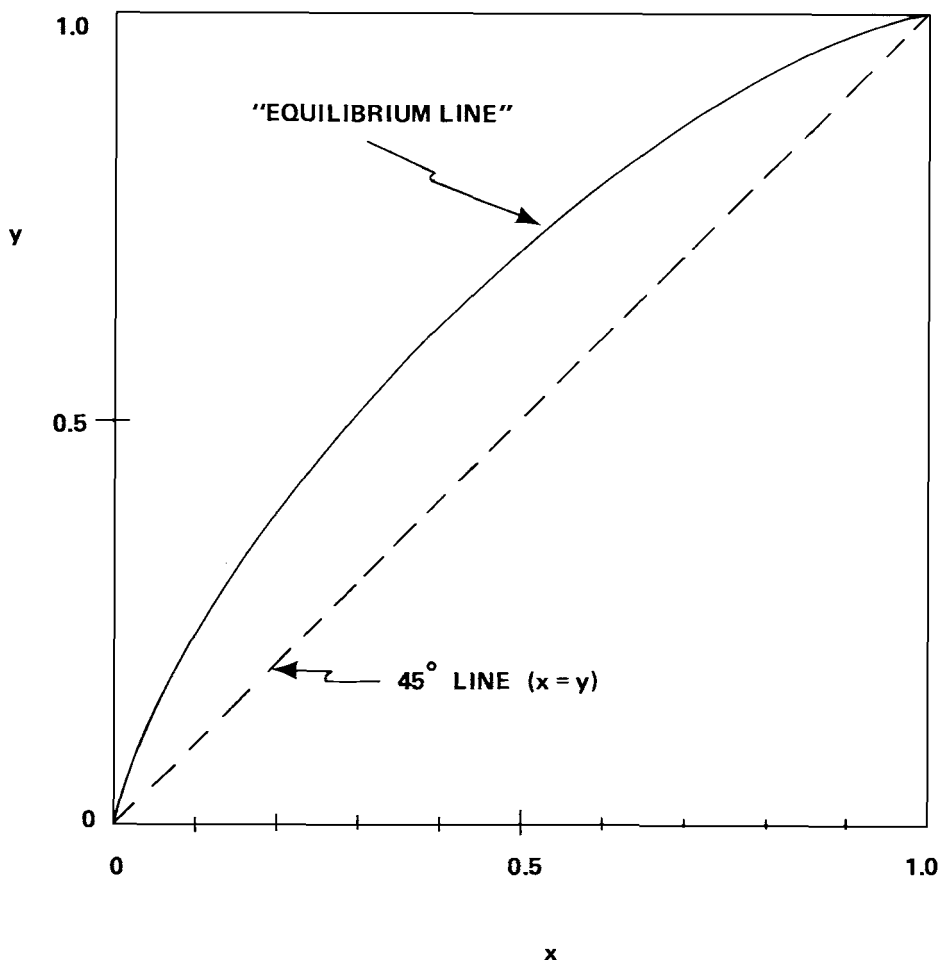


FIGURE 2.9
 x vs. y for binary mixture

The same concepts can be visualized using constant-pressure T - xy diagrams (Figure 2.11). The mixture is superheated vapor at temperatures above the dew point T_{DP} and subcooled liquid at temperatures below the bubble point T_{BP} .

Note that we can talk about either bubble-point temperature or bubble-point pressure, depending on which variable is fixed (isothermal or isobaric situations). The same is true for dew-point temperature and dew-point pressure.

VLE Calculations

Instead of using graphical techniques, we more commonly need to be able to calculate quantitatively various liquid and/or vapor compositions and temperatures or pressures given certain conditions in the column. These calculations are called bubble-point, dew-point, and flash calculations.

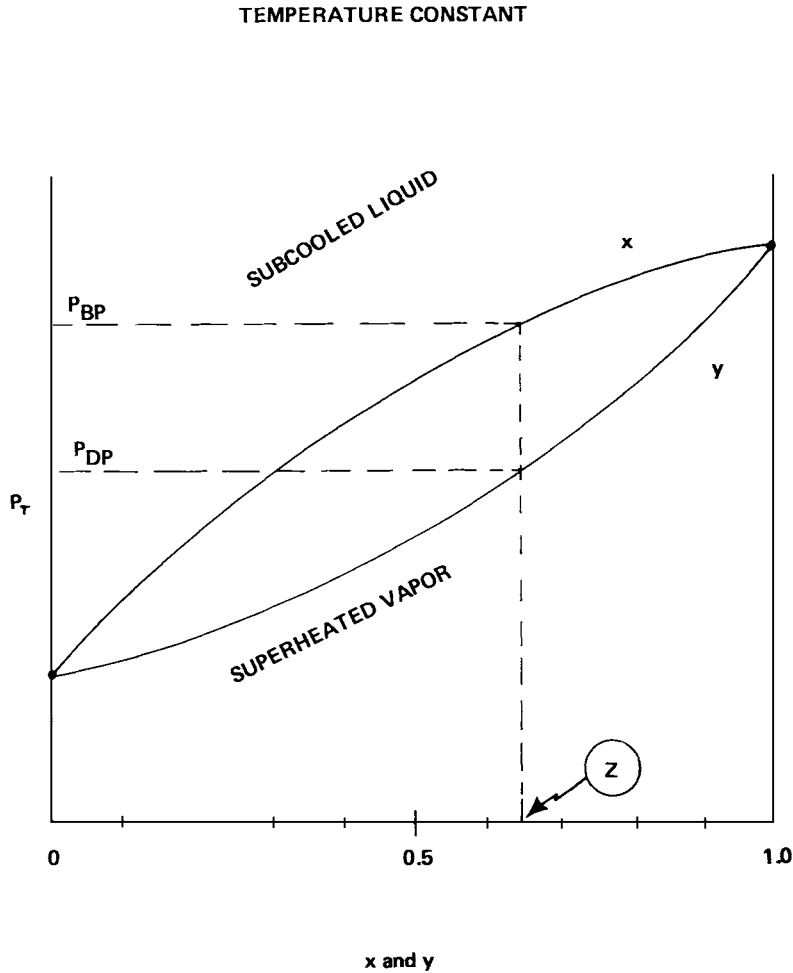


FIGURE 2.10
Bubble point and dew point at constant temperature

A. Thermodynamic Basis

The second law of thermodynamics tells us that the chemical potential of each component must be equal in both liquid and vapor phases at phase equilibrium. A somewhat simplified equation representing this condition is:

$$y_j P_T = x_j P_j \gamma_j \tag{2.4}$$

where

- x_j = mol fraction of j th component in liquid
- y_j = mol fraction of j th component in vapor
- P_T = total system pressure

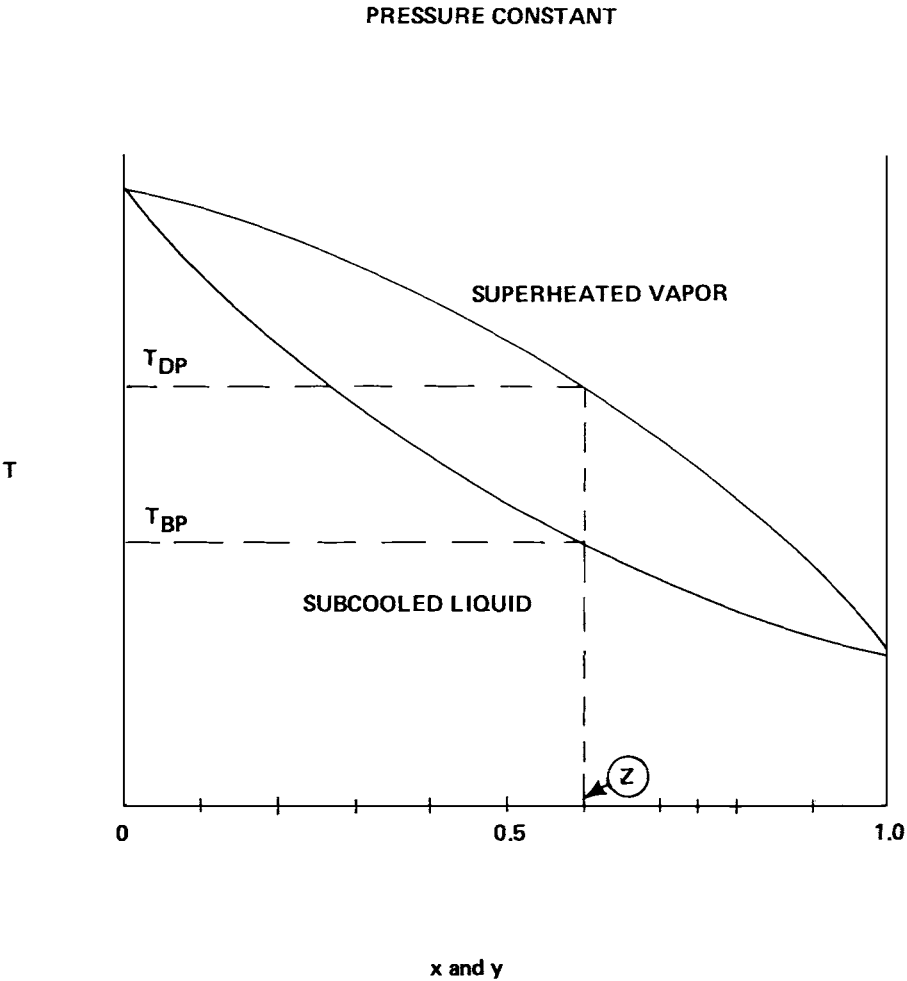


FIGURE 2.11
Bubble point and dew point at constant pressure

P_j = vapor pressure of j th component at the temperature of the system
 γ_j = activity coefficient of the j th component in the liquid phase at the conditions of temperature and composition of the liquid

This activity coefficient is a “fudge factor” that is used to account for nonideality. If the components are chemically quite similar, there is little attraction or repulsion of neighboring molecules of different types. The system is “ideal” and obeys Raoult’s law ($\gamma_j = 1$).

$$y_j P_T = x_j P_j \quad (2.5)$$

Nonideality will be discussed in more detail in Section 2.3. We will assume ideal VLE behavior for the rest of this section for purposes of simplicity.

In the petroleum industry, distribution coefficients or “ K values” are customarily used. The K value (K_j) of the j th component is defined as the ratio of vapor composition (y_j) to liquid composition (x_j).

$$K_j = \frac{y_j}{x_j} = \frac{P_j \gamma_j}{P_T}$$

If the system is “ideal” ($\gamma_j = 1$), the K value is simply the vapor pressure divided by the total system pressure.

B. Bubble-Point Calculations

In all bubble-point calculations, we know the composition of the liquid (x_j ’s are all given). In addition we must be given either the pressure or the temperature of the system. The problem is to calculate the unknown temperature or pressure and the composition of the vapor phase (y_j).

Bubble-Point Temperature Calculation. This is by far the most common type of calculation encountered in distillation work because column pressure is usually known. The calculation procedure is iterative:

1. Guess a temperature T .
2. Calculate vapor pressures of all components at T .
3. Calculate:

$$P_T^{\text{calc}}(T) = \sum_{j=1}^{N_c} x_j P_j(T) \quad (2.6)$$

where

N_c = number of components

4. Check to see if P_T^{calc} is sufficiently close to P_T .
5. If not, reguess T and go back to step 2. If P_T^{calc} is less than P , increase T . If P_T^{calc} is greater than P_T , decrease T .
6. When convergence has been achieved, calculate vapor compositions.

$$y_j = x_j P_j / P_T \quad (2.7)$$

Example. Given:

$$P_T = 1520 \text{ mm Hg}$$

$$\text{Benzene (B)} \quad x_1 = 0.40$$

$$\text{Toluene (T)} \quad x_2 = 0.35$$

$$\text{O-Xylene (X)} \quad x_3 = 0.25$$

$$\text{Guess } T = 120^\circ\text{C}$$

$$\text{Guess } T = 125^\circ\text{C}$$

	x_j	P_j at 120°C	$x_j P_j$	P_j at 125°C	$x_j P_j$	y_j
B	0.40	2300	920	2600	1040	0.671
T	0.35	1000	350	1140	399	0.258
X	0.25	380	95	440	110	0.071
			$P_T^{\text{calc}} = 1365$		$P_T^{\text{calc}} = 1549$	
			mm Hg		mm Hg	
			$P_T^{\text{calc}} < P_T$			
			T guessed is too low			
						Close enough when doing by hand

Notice the enriching of lighter component that occurs in the vapor in the above example. Benzene, the lightest component, has a higher concentration in the vapor than in the liquid. O-Xylene, on the other hand, the heaviest component, has a higher concentration in the liquid than in the vapor. This illustrates precisely why a distillation column can be used to separate chemical components. The vapor rising in the column gets richer and richer in light components at each stage. The liquid moving down the column gets richer and richer in heavy components.

Bubble-Point Pressure Calculation. In this case temperature T and liquid-phase composition are known. Total system pressure is easily calculated (with no iteration involved) from:

$$P_T = \sum_{j=1}^{N_c} x_j P_j \quad (2.8)$$

Vapor pressures P_j are known since temperature is given.

C. Dew-Point Calculations

In dew-point calculations, we know the vapor composition (y_j 's are all given) and *either* temperature *or* pressure.

Dew-Point Temperature (P Given). Solve iteratively for the temperature that satisfies equation (2.9), given P and y_j 's.

$$x_j = y_j P_T / P_j$$

$$\sum_{j=1}^{N_c} x_j = 1 = \sum_{j=1}^{N_c} (y_j / P_j) P_T$$

Rearranging:

$$P_T = \frac{1}{\sum_{j=1}^{N_c} (y_j/P_j)} \quad (2.9)$$

Dew-Point Pressure (T Given). Calculate pressure directly from equation (2.9)

D. Isothermal Flash Calculations

These calculations combine vapor-liquid equilibrium relationships with total mass and component balances. Material of known composition z_j is fed into a flash drum at a known rate of F mols/min. Both the temperature and the pressure in the drum are given. Variables that are unknown are liquid and vapor compositions and liquid and vapor flow rates. See Figure 2.12.

The equations describing the system are:

$$F = L + V \quad (2.10)$$

$$z_j F = x_j L + y_j V \quad (2.11)$$

$$y_j = x_j P_j / P_T \quad (2.12)$$

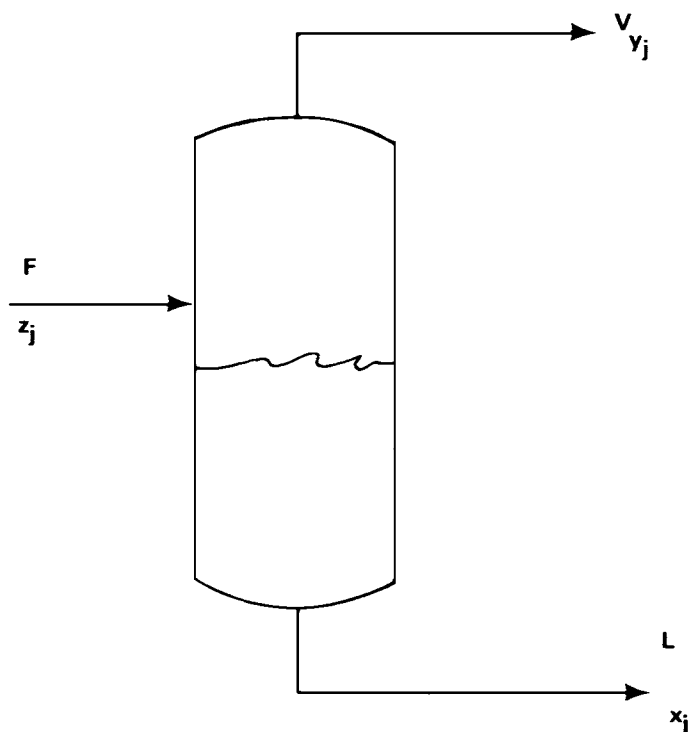


FIGURE 2.12
Isothermal flash

Combining and rearranging:

$$z_j F = x_j(F - V) + x_j P_j V / P_T$$

$$x_j = \frac{z_j}{1 + (V/F)(P_j/P_T - 1)}$$

$$\sum_{j=1}^{N_c} \frac{z_j}{1 + (V/F)(P_j/P_T - 1)} = 1 \quad (2.13)$$

The only variable that is unknown in this equation is the (V/F) ratio since the P_j 's are functions of temperature only. Therefore, another iterative, trial-and-error solution is required. One guesses a (V/F) and sees whether equation (2.13) is satisfied. The left-hand side of equation (2.13) is a nonmonotonic function (see Figure 2.13). A value of zero for (V/F) always satisfies the

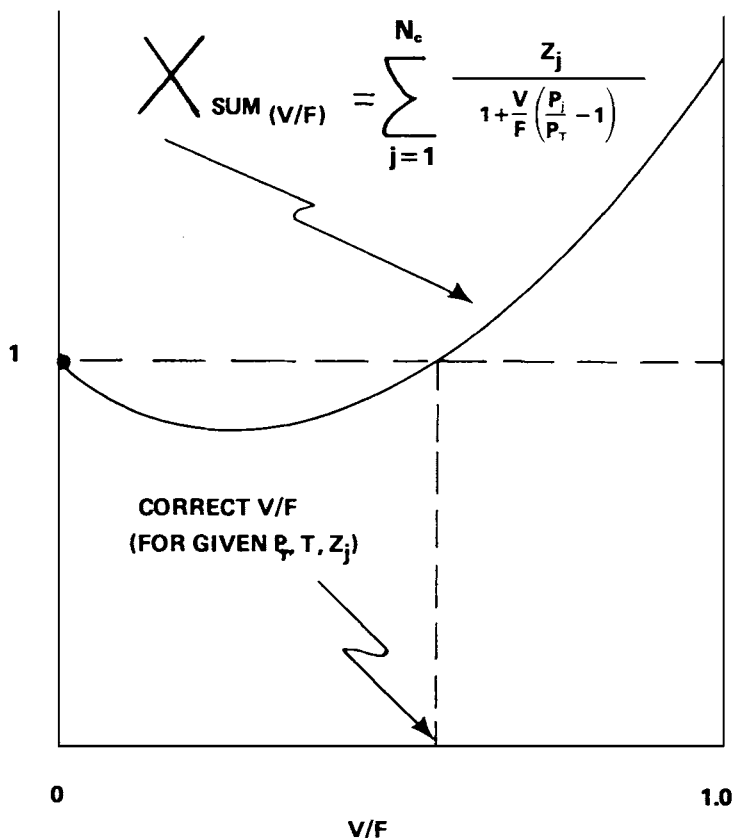


FIGURE 2.13
Graphical representation of equation (2.13)

equation, but this is not the correct, real (V/F) ratio. Before starting any flash calculation, it is vital that one checks to see that the pressure and temperature given are such that the feed mixture is in the two-phase region. That is, the system pressure must be between the bubble-point pressure and the dew-point pressure for a mixture with a composition equal to the feed composition and at the given temperature T .

$$P_{DP} < P_T < P_{BP} \quad (2.14)$$

$$P_{DP} = \frac{1}{\sum_{j=1}^{N_c} z_j/P_j(T)} \quad (2.15)$$

$$P_{BP} = \sum_{j=1}^{N_c} z_j P_j(T) \quad (2.16)$$

Since T is known, both P_{DP} and P_{BP} can be calculated explicitly with no iteration involved.

Relative Volatility

Relative volatility is a very convenient measure of the ease or difficulty of separation in distillation. The volatility of component j relative to component k is defined as:

$$\alpha_{jk} = \frac{y_j/x_j}{y_k/x_k} \quad (2.17)$$

A large value of relative volatility α_{jk} implies that components j and k can be easily separated in a distillation column. Values of α_{jk} close to 1 imply that the separation will be very difficult, requiring a large number of trays and high energy consumption.

For binary systems relative volatility of light to heavy component is simply called α :

$$\alpha_{LH} = \alpha = \frac{y/x}{(1-y)/(1-x)} \quad (2.18)$$

where x and y are mol fractions of light component in the liquid and vapor phases respectively. Rearrangement of equation (2.18) leads to the very useful y - x relationship that can be employed when α is constant in a binary system.

Figure 2.14 sketches y versus x lines for various values of α . The larger the relative volatility α , the fatter is the equilibrium curve. For an ideal (Raoult's

law) binary system, α can be expressed very simply as the ratio of the vapor pressures of light and heavy components.

$$yP_T = P_L x$$

$$(1 - y)P_T = P_H(1 - x) \quad (2.19)$$

$$\alpha = \frac{y/x}{(1 - y)/(1 - x)} = \frac{P_L}{P_H} \quad (2.20)$$

If the temperature dependence of the vapor pressure of both components is the same, α will be independent of temperature. In other words, relative volatility is constant if the vapor pressure lines are parallel in a $\ln P$ versus $1/T$ plot. This is true for many components over a limited temperature range, particularly when the components are chemically similar. Distillation columns are frequently designed assuming constant relative volatility because it greatly simplifies the vapor–liquid equilibrium calculations. Relative volatilities usually decrease somewhat with increasing temperature in most systems.

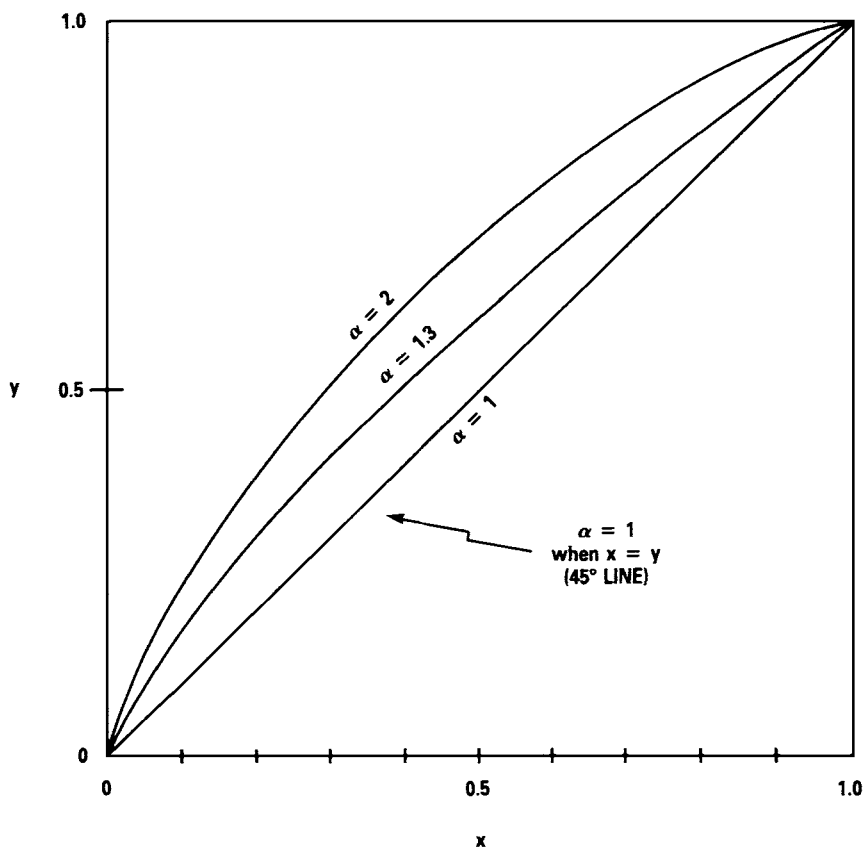


FIGURE 2.14
Relative volatility on x-y diagram

For multicomponent systems, applying the basic definition [equation (2.17)] and rearranging lead to the following explicit relationship between any vapor composition y_j and given liquid compositions (x_j 's) and relative volatilities (α_j 's).

$$y_j = \frac{\alpha_j x_j}{\sum_{k=1}^N \alpha_k x_k} \quad (2.21)$$

where α_j is the volatility of the j th component relative to some arbitrary base component (usually chosen as the heaviest component.)

Example. Given the liquid compositions and relative volatilities, calculate the vapor compositions:

x_j	α_j	$x_j \alpha_j$	y_j
0.35	3.5	1.225	0.517
0.45	2.1	0.945	0.399
0.20	1	0.200	0.084

Nonideality

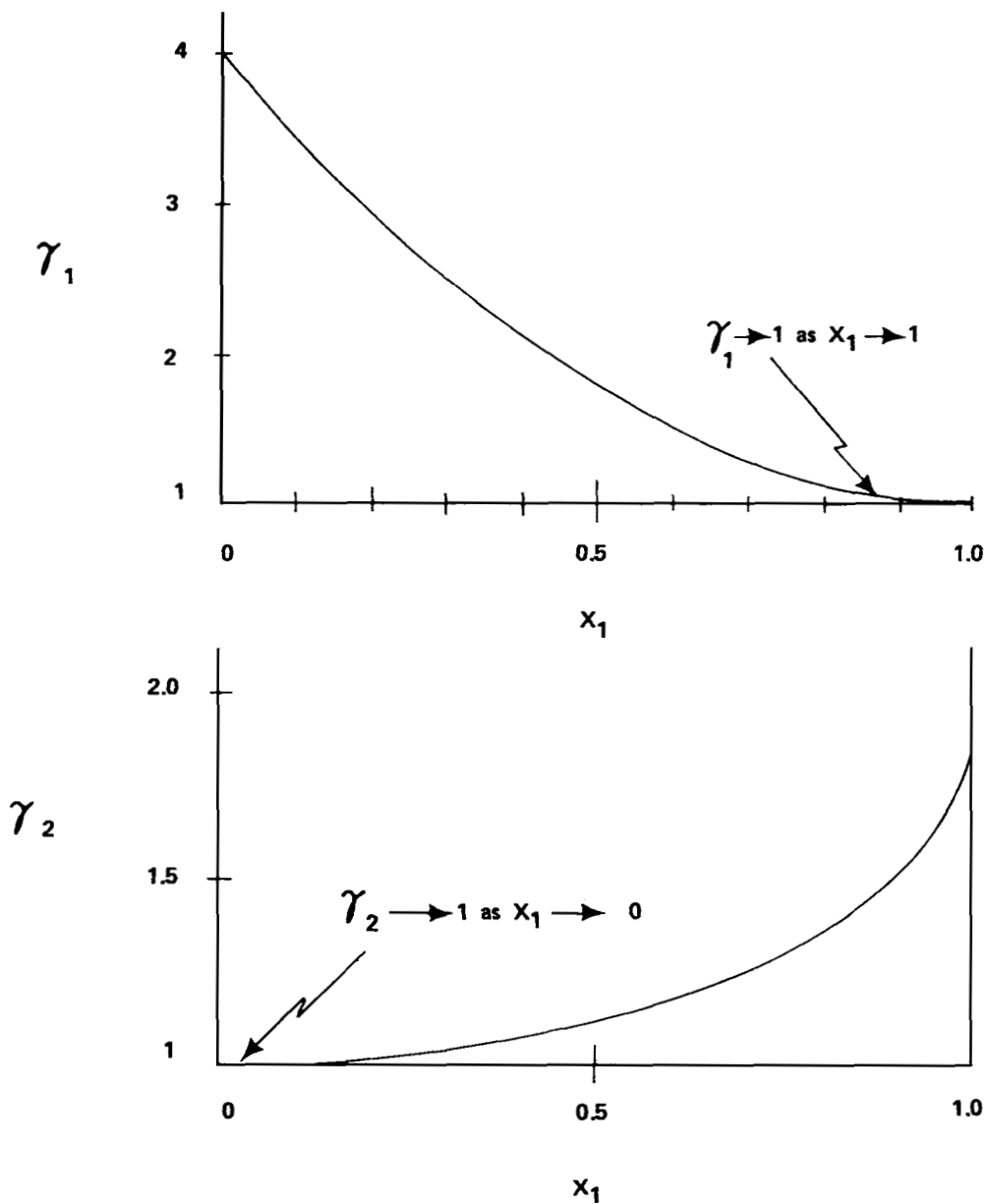
In most distillation systems, the predominant nonideality occurs in the liquid phase because of molecular interactions. Equation (2.4) contains γ_j , the liquid-phase activity coefficient of the j th component.

When chemically dissimilar components are mixed together (for example, oil molecules and water molecules), there can be repulsion or attraction between dissimilar molecules. If the molecules repel each other, they exert a higher partial pressure than if they were ideal. In this case the activity coefficients are greater than unity (called a "positive deviation" from Raoult's law). If the molecules attract each other, they exert a lower partial pressure than if they were ideal. Activity coefficients are less than unity (negative deviations).

Activity coefficients are usually calculated from experimental data. Figure 2.15 sketches typical activity coefficient data as a function of the light component composition x_1 . Empirical equations (Van Laar, Margules, Wilson, etc.) are used to correlate activity coefficient data.

Azeotropes occur in a number of nonideal systems. An azeotrope exists when the liquid and vapor compositions are the same ($x_j = y_j$). There are several types of azeotropes. Figures 2.16, 2.17, and 2.18 sketch typical phase diagrams for these.

Negative deviations (attraction) can give a higher temperature boiling mixture than the boiling point of the heavier component. This can lead to formation of a "maximum-boiling" azeotrope (Figure 2.16). Positive deviations (repulsion) can give a lower temperature boiling mixture than the boiling point of the light component. A modest amount of repulsion can lead to the formation of a minimum boiling azeotrope (Figure 2.17). If the repulsion is very strong, the system may break into two-liquid phases with different compositions in each liquid phase. This is called a "heterogeneous" azeotrope (Figure 2.18).

**FIGURE 2.15****Typical activity coefficients as functions of light component composition**

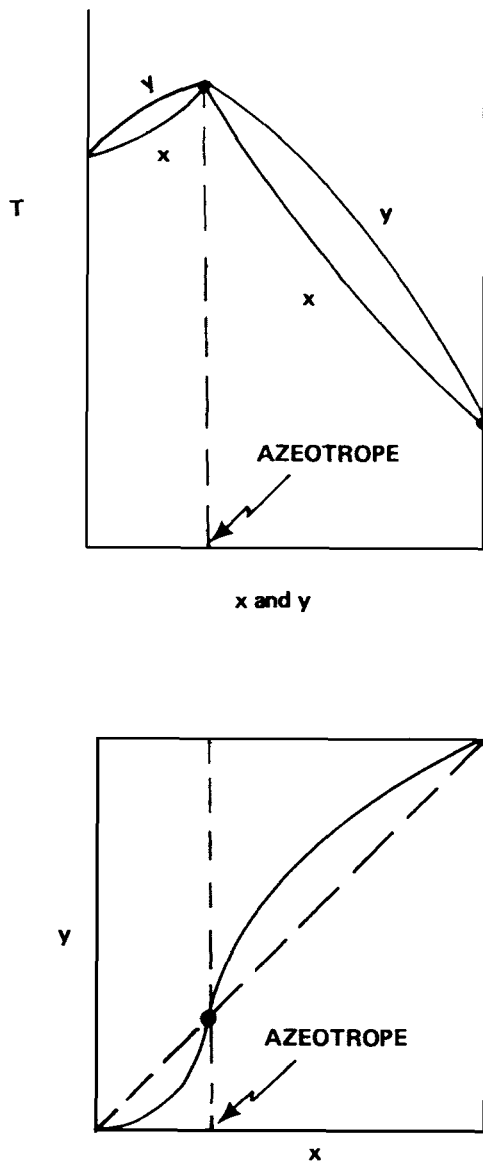


FIGURE 2.16
Typical homogeneous “maximum boiling” azeotrope

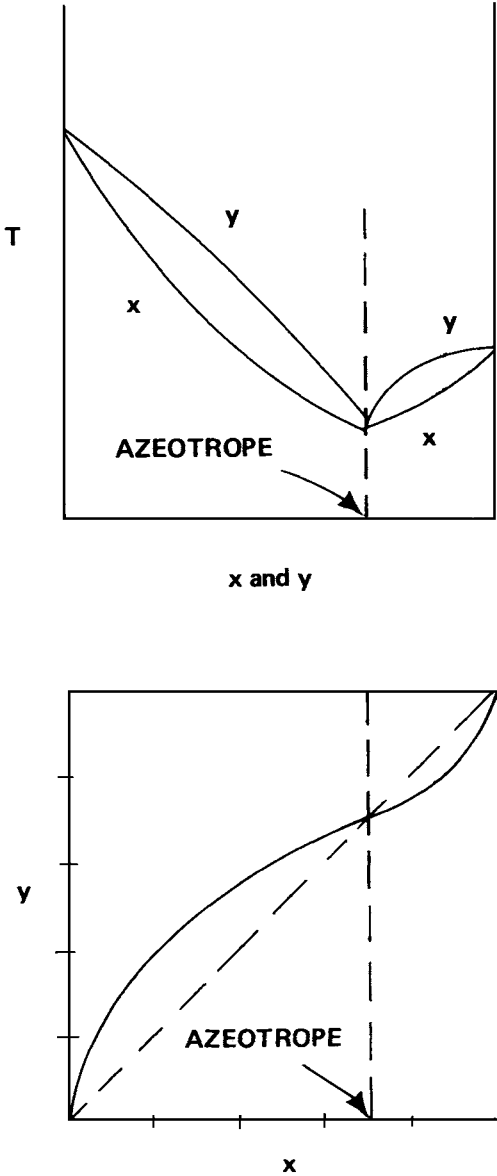


FIGURE 2.17
Homogeneous “minimum boiling” azeotrope

A detailed discussion of azeotropes is beyond the scope of this brief introduction. The control engineer should be aware that the existence of azeotropes imposes restrictions on the operation and performance of a distillation column.

2.4 GRAPHICAL SOLUTION TECHNIQUES

The equations describing a binary distillation column can be solved graphically using the famous McCabe–Thiele diagrams. These techniques are very useful in gaining an appreciation of the effects of various design and operating parameters.

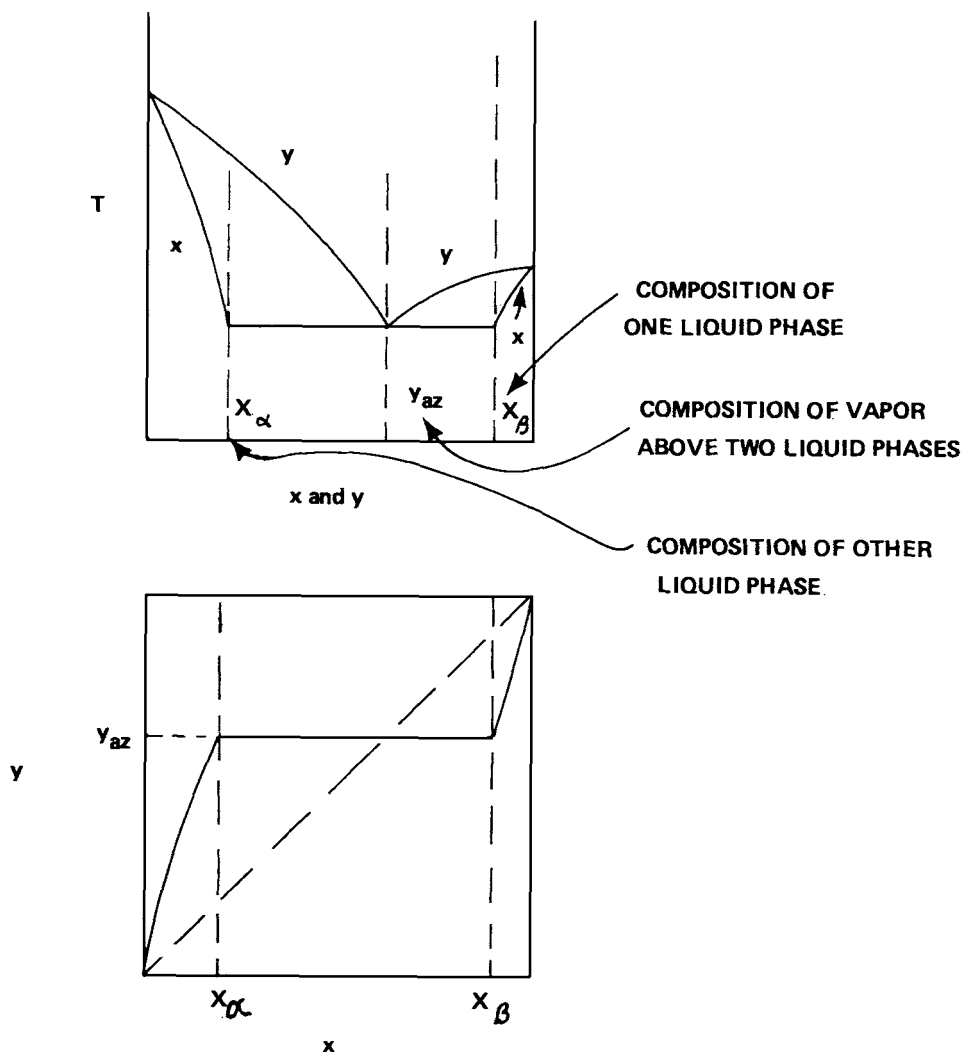


FIGURE 2.18
Heterogeneous azeotropes

These effects are usually apparent on the diagrams, so one gets a picture of the process.

System

Figure 2.19 summarizes the system and nomenclature to be considered. Liquid and vapor rates in the section above the feed ("rectifying" section) are called L_R and V_R in mols/minute. Liquid and vapor rates below the feed in the stripping section are called L_S and V_S . These are assumed to be constant

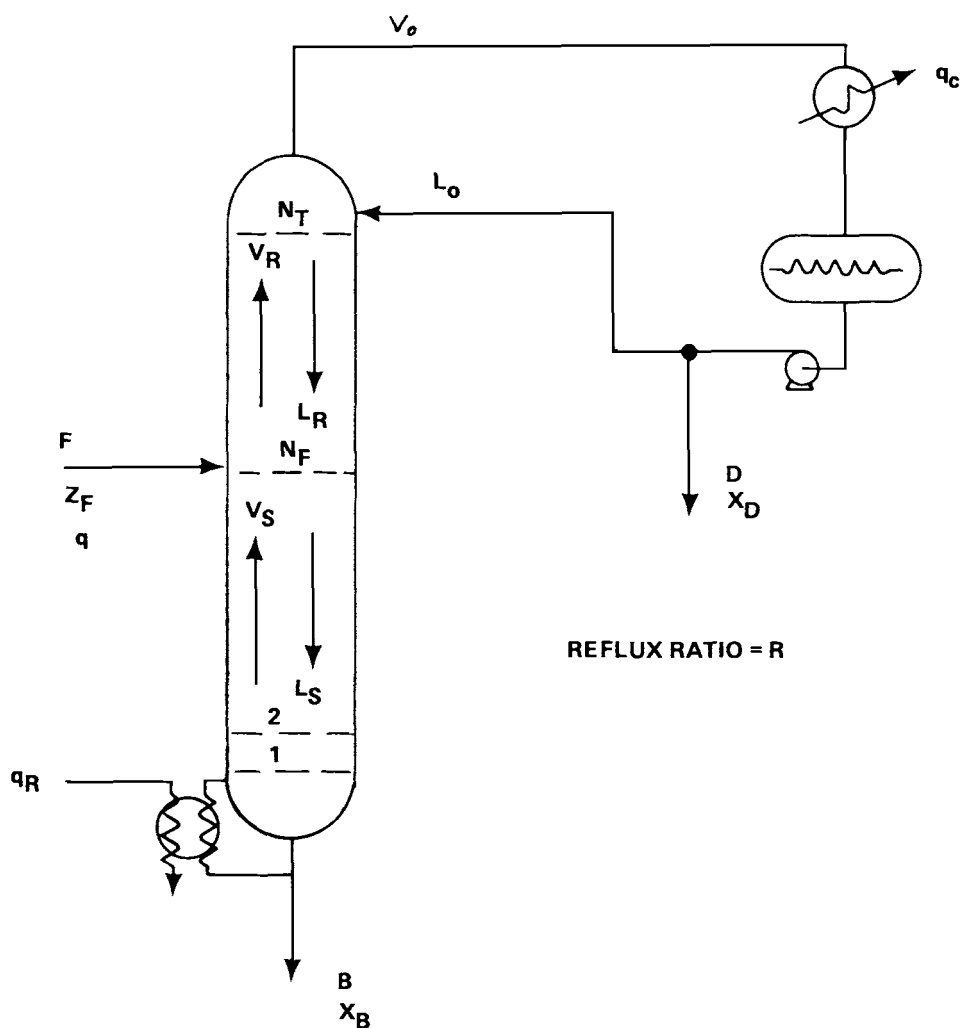


FIGURE 2.19
Simple distillation column

on all trays throughout each individual section. This constancy of molar flow rates is what we call “equimolal overflow.” Primarily it assumes that the molar heats of vaporization of the components are about equal. In many systems this is a pretty good assumption.

A total condenser is used to produce liquid reflux and distillate product. The reboiler is a “partial reboiler” (vapor is boiled off a liquid pool). The composition of this liquid pool is the same as the bottom product composition. Thermosyphon, kettle, internal, and forced-circulation reboilers are all usually partial reboilers.

The parameter q will be used to describe the thermal condition of the feed. The ratio of the internal reflux flow rate L_R to the distillate flow rate D is called the reflux ratio R .

$$R = L_R/D \quad (2.22)$$

Reflux ratio is widely used as an indication of the energy consumption.

Equations

A. Overall Balances

Mass and component balances can be written around the entire column system.

$$F = D + B \quad (2.23)$$

$$Fz_F = x_D D + x_B B \quad (2.24)$$

These relationships must be satisfied under steady-state conditions. Note that distillate and bottom product rates can be calculated from equations (2.23) and (2.24) if feed conditions and product compositions are specified.

$$D = F(z_F - x_B)/(x_D - x_B) \quad (2.25)$$

$$B = F - D \quad (2.26)$$

B. Stripping Section

A light component balance around the n th tray in the stripping section yields (see Figure 2.20):

$$L_S x_{n+1} = V_S y_n + B x_B \quad (2.27)$$

Rearranging:

$$y_n = \left(\frac{L_S}{V_S} \right) x_{n+1} + \left(\frac{-B}{V_S} \right) x_B \quad (2.28)$$

This is called the operating-line equation. It has the form of a straight line: $y = mx + b$. The slope of the line is the ratio of liquid to vapor flow rates in the stripping section. This straight line can be plotted on an x - y diagram (see Figure 2.21).

The intersection of the operating line with the 45° line (where $x = y$) occurs at x_B . This is easily proved by letting the point of intersection be $x_i = y_i$. Substituting into equation (2.27) gives:

$$\begin{aligned} L_S x_i &= V_S y_i + B x_B \\ L_S x_i &= V_S x_i + B x_B \\ (L_S - V_S)x_i &= B x_B \end{aligned} \quad (2.29)$$

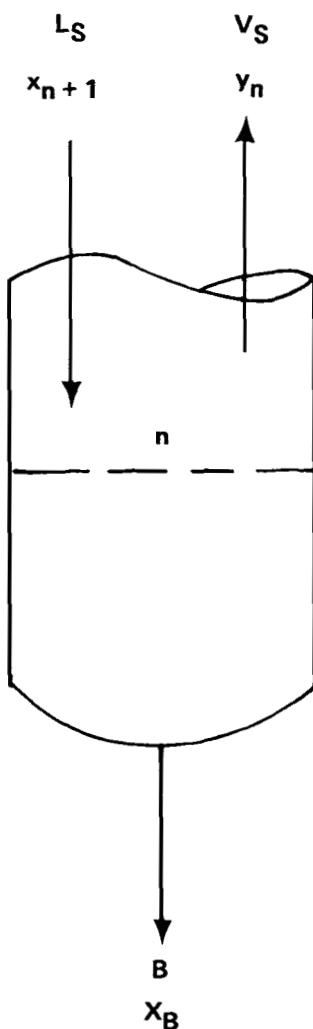


FIGURE 2.20
Material balance on stripping section

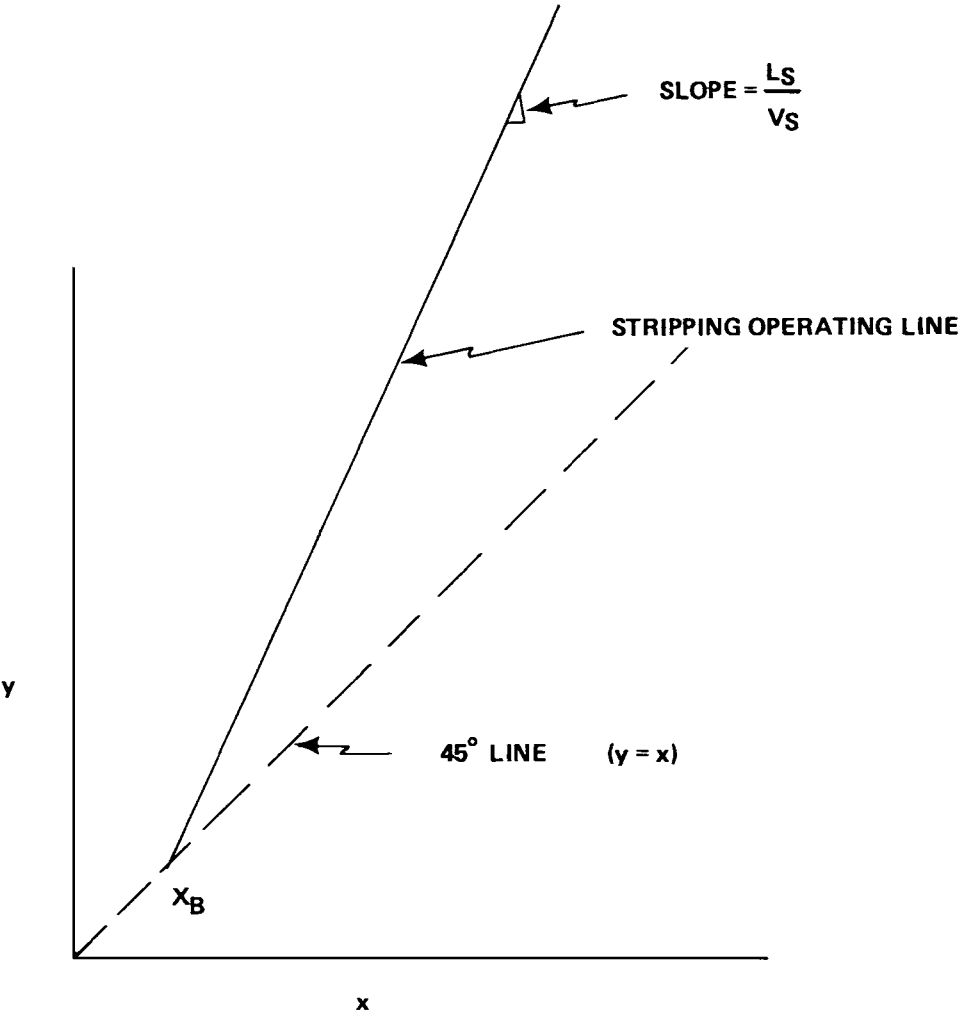


FIGURE 2.21
Operating line of stripping section

From a total mass balance around the system in Figure 2.20, $L_S = V_S + B$. Therefore, $L_S - V_S = B$. Substituting into equation (2.29) gives:

$$B x_i = B x_B$$

$$x_i = x_B$$

Therefore, the operating line can be easily placed on an x - y diagram by simply drawing a straight line with slope (L_S/V_S) , starting at x_B on the 45° line (see Figure 2.21).

C. Rectifying Section

A similar component balance around the upper part of the column above the n th tray in the rectifying section (see Figure 2.22) gives:

$$V_R y_n = L_R x_{n+1} + D x_D$$

$$y_n = \frac{L_R}{V_R} x_{n+1} + \frac{D}{V_R} x_D \quad (2.30)$$

This is the equation of another straight line called the rectifying operating line.

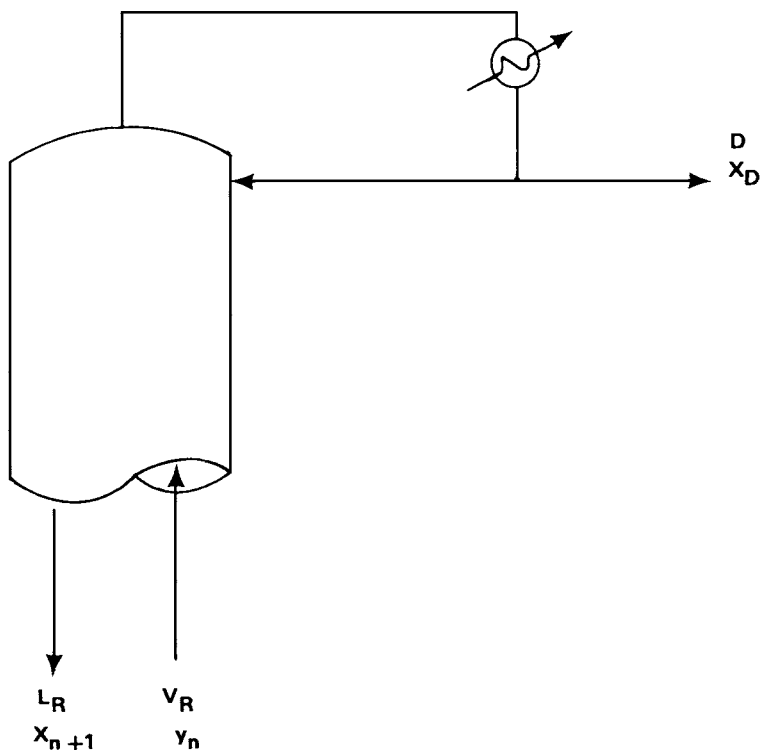


FIGURE 2.22
Material balance on rectifying section

It intersects the 45° line at x_D and has a slope equal to the ratio of the liquid-to-vapor rates in the rectifying section. Figure 2.23 shows both operating lines together with the y - x vapor–liquid equilibrium curve.

Stepping Off Trays

Tray-to-tray calculations involve the solution of vapor–liquid equilibrium relationships and component balances. These tray-to-tray calculations can be solved graphically by stepping back and forth between the operating line ($y = mx + b$) and the VLE curve ($y = f(x)$), as shown in Figure 2.24.

For example, let us start at the bottom of the column with known values of x_B and L_S/V_S . Our VLE relationship gives us y_B if we know x_B . Graphically,

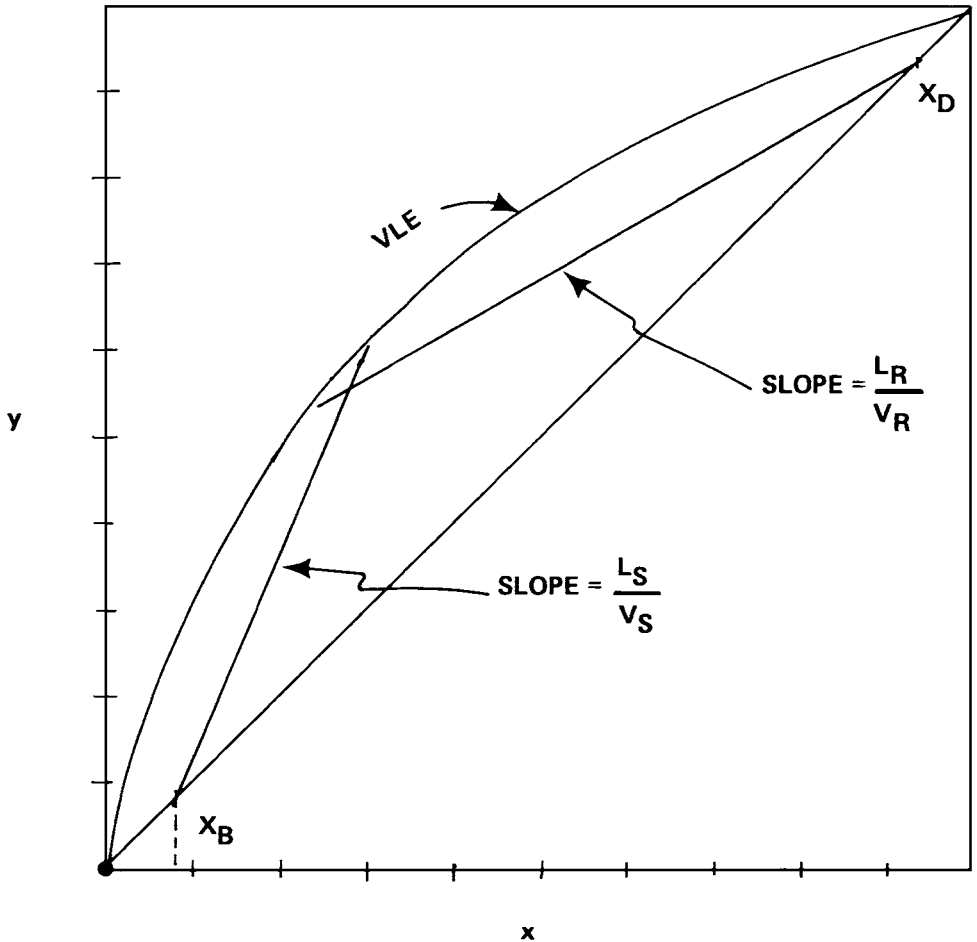


FIGURE 2.23

x - y diagram showing both stripping and rectifying operating lines

we simply move vertically up a line through x_B until the VLE curve is intersected, and then read off the y_B value.

A component balance around the reboiler gives:

$$L_S x_1 = V_S y_B + B x_B \quad (2.31)$$

$$y_B = \frac{L_S}{V_S} x_1 + \left(\frac{-B}{V_S} \right) x_B \quad (2.32)$$

This is the stripping operating line [equation (2.28)] applied to the reboiler stage. We know y_B . We could plug into equation (2.31) and solve analytically for x_1 . Alternatively, we can solve graphically for x_1 simply by moving horizontally on a straight line through y_B until the operating line is intersected. Then this

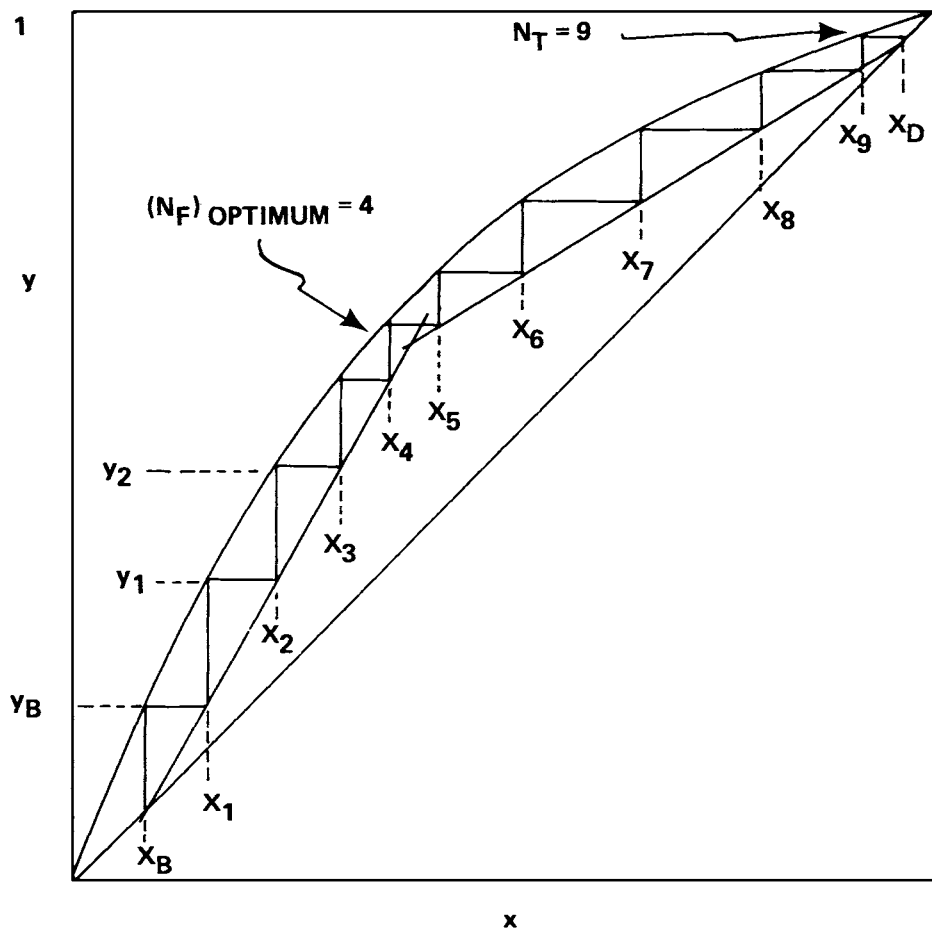


FIGURE 2.24
McCabe-Thiele diagram—stepping between VLE curve and operating lines to estimate number of trays required

same procedure is applied again to the first tray. y_1 is determined from x_1 (move vertically to the VLE curve). x_2 is determined from y_1 by moving horizontally to the stripping operating line.

This stepping procedure is continued up through the stripping section until the intersection of the operating lines is passed (see Figure 2.24). This determines the number of trays required in the stripping section. Then the rectifying operating line is used, and the stepping is continued until x_D is reached. The number of trays in the rectifying section can be determined in this manner. Thus we can design a column (i.e., calculate the total number of trays N_T and the feed tray N_F) using this graphical technique, having specified product compositions and operating line slopes. As we will show in the next section, the operating line slopes are both known if the feed thermal condition and reflux ratio have been specified.

Feed Thermal Condition

The feed to a distillation column can be liquid or vapor, or both, depending on the temperature, pressure, and composition of the feed. To quantify the thermal condition of the feed, the parameter q is defined as the fraction of the feed that is liquid.

$$q = \frac{L_S - L_R}{F} \quad (2.33)$$

It follows that $1 - q$ is the fraction of feed that is vapor:

$$1 - q = \frac{V_R - V_S}{F} \quad (2.34)$$

If the feed is a saturated liquid at its bubble point, $q = 1$. If the feed is a saturated vapor at its dew point, $q = 0$. If the feed is a vapor-liquid mixture, q is a fraction. Values of q greater than 1 indicate subcooled liquid feed. Values of q less than 0 indicate superheated vapor feed.

Now let us look at the intersection points (x_i, y_i) of the stripping and rectifying operating lines. Use of equations (2.28) and (2.30) gives:

$$V_R y_i = L_R x_i + x_D D$$

$$V_S y_i = L_S x_i - x_B B$$

Subtracting gives:

$$(V_R - V_S)y_i = (L_R - L_S)x_i + (x_D D + x_B B)$$

Using equations (2.24), (2.33) and (2.34) gives:

$$\begin{aligned} F(1 - q)y_i &= (-qF)x_i + z_F F \\ y_i &= \left(\frac{-q}{1 - q} \right) x_i + \left(\frac{-z_F}{1 - q} \right) \end{aligned} \quad (2.35)$$

This is the equation of a straight line with slope $(-q/(1 - q))$. It is called the q line and intersects the 45° line at z_F . Thus the intersection of the operating lines must lie on the q line, which can be easily drawn given z_F and q . Figure 2.25 shows q lines for several values of q .

The slope of the rectifying operating line (L_R/V_R) can be expressed in terms of the internal reflux ratio $L_R/D = R$. If the reflux is a saturated liquid at its bubble point,* $L_R = L_o$.

$$\begin{aligned}\frac{L_R}{V_R} &= \frac{L_o}{V_R} = \frac{L_o}{L_o + D} = \frac{L_o/D}{L_o/D + 1} \\ \frac{L_R}{V_R} &= \frac{R}{R + 1}\end{aligned}\quad (2.36)$$

Thus the rectifying operating line can be drawn if x_D and R are specified. It is a straight line with a slope of $R/(R + 1)$, intersecting the 45° line at x_D . Then the stripping operating line can be drawn if x_B and q are specified. It is a straight line joining x_B on the 45° line with the intersection of the rectifying operating line and the q line.

Design Problem

We are now ready to summarize the graphical design technique for determining the number of trays required to achieve desired product purities, given a reflux ratio. As we will show later, the lower the reflux ratio specified, the more trays are required. Since increasing reflux ratio increases energy costs ($V_o = L_o + D$), while increasing trays increases investment costs, distillation design involves a classical engineering trade-off between the two design variables: reflux ratio and number of trays. This will be discussed further under "Limiting Conditions."

Given:

1. Feed: z_F , F , q
2. Desired product purities: x_D , x_B
3. VLE curve in x - y coordinates
4. Operating external reflux ratio: $R = L_o/D$

* If external reflux is subcooled—usually the case—then

$$L_R = L_o \left[1 + \frac{C_{pm}}{\lambda_M} (T_o - T_R) \right]$$

where C_{pm} = molar specific heat, λ_M = molar latent heat, T_o = vapor temperature, and T_R = temperature of L_o . Then

$$V_o = V_R - L_o \left[\frac{C_{pm}}{\lambda_M} (T_o - T_R) \right].$$

Calculate:

1. Total number of trays: N_T
2. Feed tray location: N_F

Procedure:

1. Draw VLE curve.
2. Draw 45° line.
3. Locate x_B , x_D , and z_F on 45° line.
4. Calculate slope of q line ($-q/(1 - q)$) and draw q line from z_F point on 45° line.
5. Calculate B and D from overall balances [equations (2.25) and (2.26)].
6. Calculate liquid and vapor flow rates in rectifying and stripping sections.

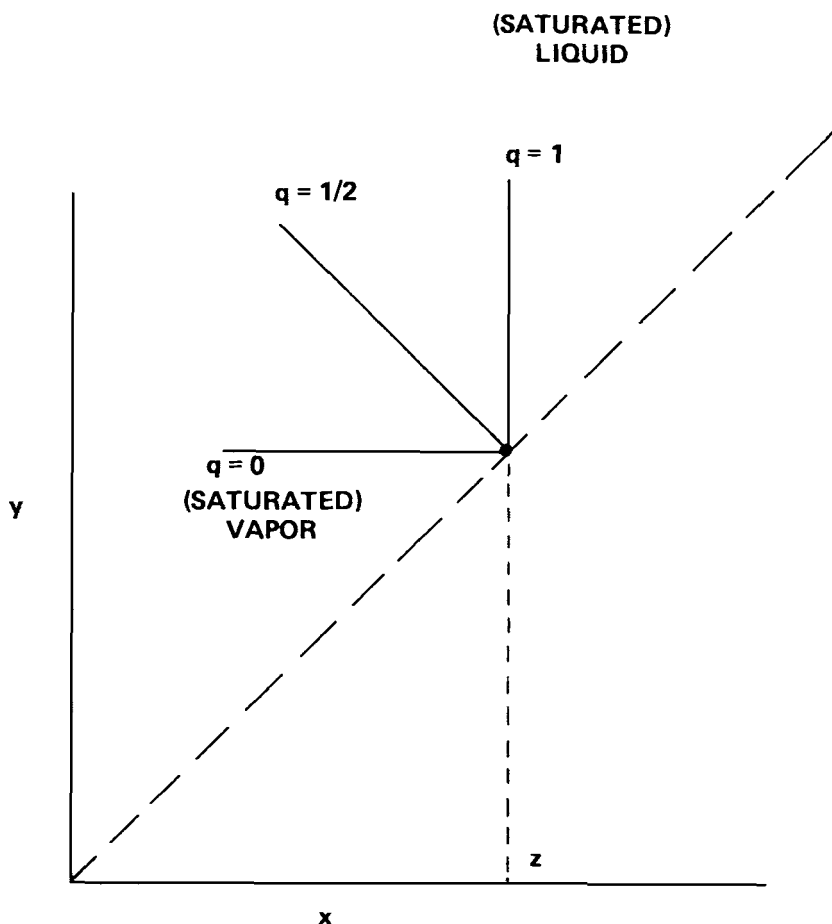


FIGURE 2.25
q-line on x-y diagram

If the external reflux, L_o , is at its bubble point:

$$L_R = L_o = (R) D \quad (2.37)$$

$$V_R = L_R + D \quad (2.38)$$

$$V_S = V_R - (1 - q) F \quad (2.39)$$

$$L_S = L_R + qF \quad (2.40)$$

7. Calculate slopes of operating lines: rectifying = L_R/V_R ; stripping = L_S/V_S .

8. Draw rectifying operating line from x_D point on 45° line with slope L_R/V_R .

9. Draw stripping operating line from x_B point on 45° line to the intersection of the q line with the rectifying operating line.

10. Check your calculations by seeing whether the slope of stripping operating line is L_S/V_S .

11. Start from the x_B point on the 45° line and step up the column from the stripping operating line to the VLE curve. The first step corresponds to the partial reboiler. The next step is tray 1, the next is tray 2, and so on.

12. When this stepping procedure crosses the intersection of operating lines, this is the “optimum” feed tray (i.e., any other feed tray would require a greater total number of trays). Thus the feed tray N_F has been determined.

13. Switch to the rectifying operating line and continue stepping.

14. When the x_D point is crossed, this is the total number of trays N_T .

This last step will not go through the x_D point exactly, implying a noninteger number of trays.

Don't let this worry you. In this procedure we have assumed “perfect” or “theoretical” or “100 percent efficient” trays; that is, trays on which the vapor and liquid streams leaving the trays are in perfect phase equilibrium. Actual industrial columns seldom achieve this ideal situation, so an efficiency factor must be used to determine the number of actual trays installed in the column (which must be an integer number). Typical efficiencies run from 40 to 90 percent, depending on the system.

Rating Problems

The graphical McCabe–Thiele methods studied in the previous sections for the design of distillation columns are also widely used to analyze the operation of an existing column. In this case the total number of trays in the column N_T is fixed. The feed tray may also be fixed, or, if there are multiple feed points available on the column, it may be varied. These fixed-column problems are called “rating problems,” as opposed to the “design problems,” in which N_T is calculated.

There are a variety of possible rating problems. The two most commonly encountered are (with N_T and N_F fixed):

—To determine the reflux ratio required to achieve specified product purities x_B and x_D

—To determine the product compositions that result from specified values of reflux ratio and distillate flow rate

Both of these calculations involve iterative, trial-and-error solution techniques. Basically one guesses a solution and sees if the stepping procedure produces *exactly* the same number of trays in each section as has been specified (see Figure 2.26).

Notice that in both of these problems, two variables must be specified to define the system completely. This magic number of *two* occurs again and again in distillation (see Section 4). It is often called the “degrees of freedom” of the system. Mathematically the two degrees of freedom are the result of subtracting all the constraining equations describing the system (mass, component, and energy balances; VLE equilibrium relationships; and specified variables) from the total number of system variables.

Limiting Conditions

McCabe–Thiele diagrams are useful for getting a clear picture of some of the limiting conditions on the separation that can be achieved in a distillation column.

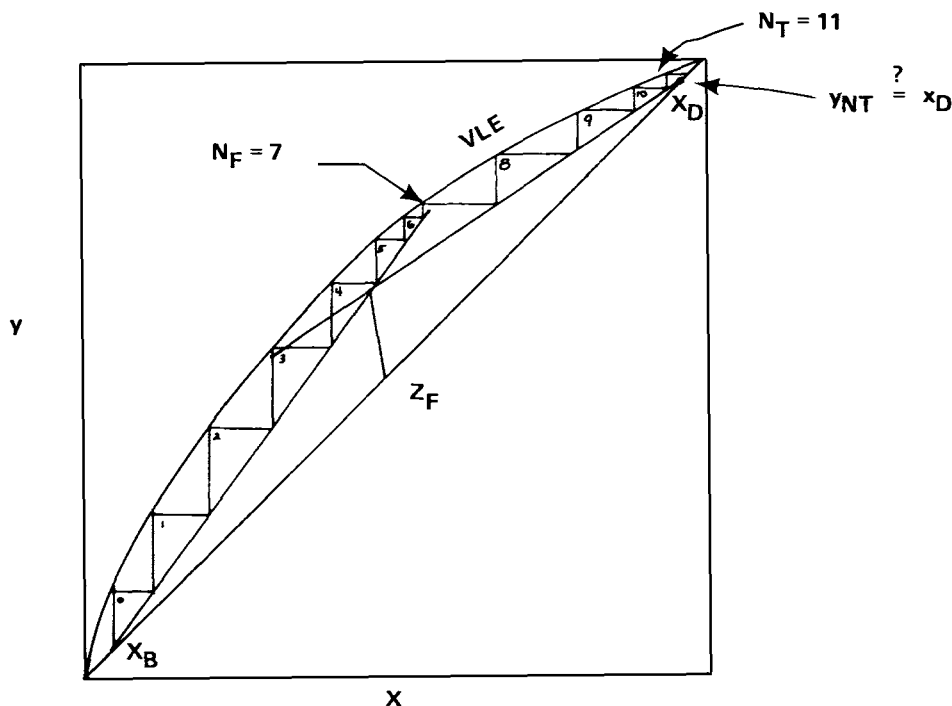


FIGURE 2.26
McCabe–Thiele diagram for rating problem

A. Minimum Reflux Ratio

The minimum reflux ratio (for specified product purities and feed conditions) occurs when an infinite number of trays are required to make the separation. Figure 2.27 shows the normal minimum reflux ratio situation. It occurs when the operating lines just intersect on the VLE curve. An infinite number of trays are required to step past the feed plate because of the “pinch” condition (the converging operating and VLE lines). The actual reflux ratio used must be higher than the minimum.

Increasing reflux ratio requires fewer trays (less capital cost) but increases energy costs. Economic optimization studies have led to the commonly used heuristic (rule of thumb) that the optimum actual reflux ratio is 1.1 to 1.2 times the minimum reflux ratio (see Figure 2.28).

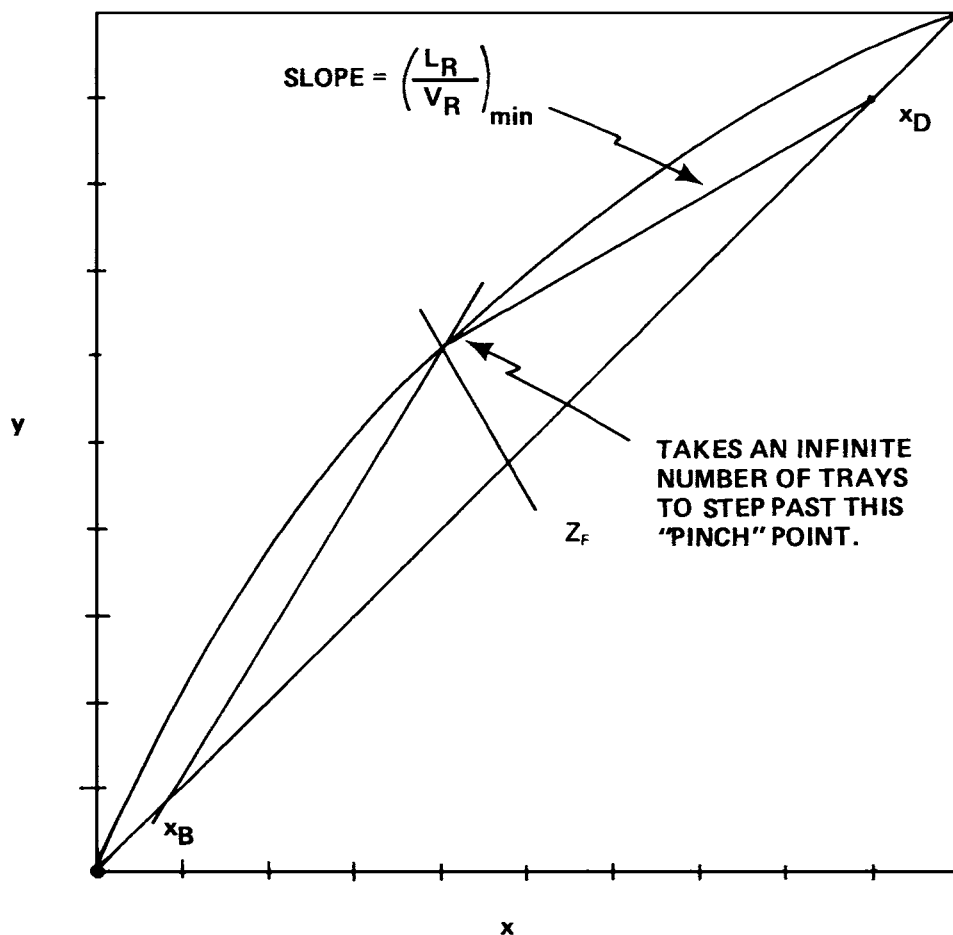


FIGURE 2.27
Column operation at minimum reflux ratio

In some unusual VLE systems, the pinch between the VLE curve and an operating line can occur at some point other than the feed point.

B. Minimum Number of Trays

The minimum number of trays to make a specified separation is found when an infinitely large reflux ratio is used. The L/V ratios in both sections of the column become unity and lie on the 45° line (Figure 2.29).

This situation actually takes place in a column when it is operated under “total reflux” conditions. No feed is introduced and no products are withdrawn, but heat is added in the reboiler and all the overhead vapor is condensed and returned to the column as liquid reflux. Since $D = 0$, $R/D = \infty$. Also:

$$\frac{L_R}{V_R} = \frac{L_S}{V_S} = 1$$

Thus a column with fewer than the minimum number of trays cannot achieve the desired separation, even at very high reflux ratios.

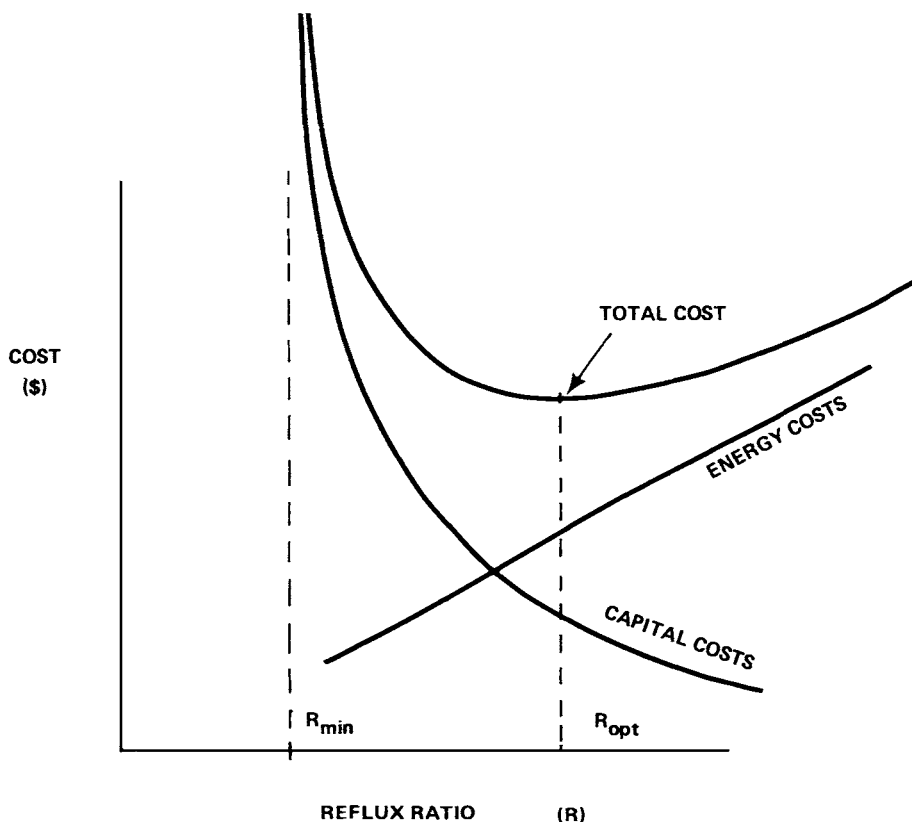


FIGURE 2.28
Costs vs. reflux ratio

For a system with constant relative volatility α_{LH} , the Fenske equation can be used to solve analytically for the minimum number of trays $(N_T)_{\min}$.

$$(N_T)_{\min} + 1 = \frac{\log \frac{x_{DL}/x_{DH}}{x_{BL}/x_{BH}}}{\log \alpha_{LH}}$$

A partial reboiler is assumed in the above equation. The L and H subscripts refer to light and heavy components.

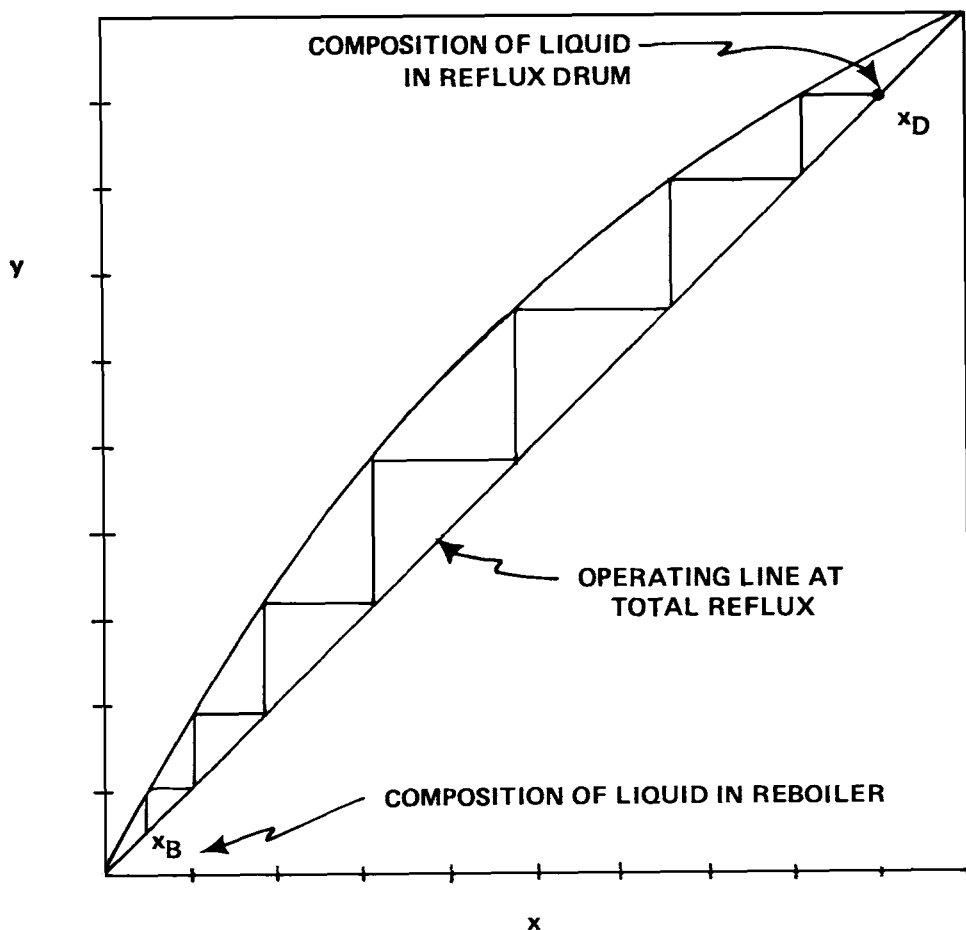


FIGURE 2.29
Minimum number of trays required at total reflux

2.5 EFFECTS OF VARIABLES

Now that the McCabe–Thiele method has been introduced, we can visualize the effects of various operating and design parameters.

Design Case

A. Increasing Product Purities (Raising x_D and/or Decreasing x_B)

The number of trays required in the column is increased. The minimum reflux ratio is increased somewhat, but this increase is only very gradual since the upper end of the rectifying operating line merely approaches closer and closer to the (1, 1) point, which changes the operating line slope only slightly at high purities.

B. Increasing Relative Volatility

The number of trays is reduced. Minimum reflux ratio is also reduced. Relative volatility has a strong effect on the cost of separation. Therefore, columns are usually designed to operate at the lowest economical pressure, since lower pressure means lower temperatures and higher relative volatilities in most systems. The lowest economical pressure is usually the pressure that provides a temperature in the overhead condenser that is high enough to permit cooling water or air to be used for heat removal. Going to lower pressures would require refrigeration, which is very expensive.

C. Increasing Feed q

Reflux ratio is reduced as the feed is made colder (as q increases) but energy input to the reboiler increases.

D. Changing Feed Composition z_F

Thermodynamics tells us that the maximum work of separation occurs when the feed is a 50/50 mixture. Therefore, higher or lower feed compositions should require less energy.

If two or more feed streams with different compositions are to be separated in the column, they should not be mixed and fed in at a single feed point. Instead they should be fed on separate feed trays at locations where tray compositions approximate feed compositions.

Operating Case (N_T and N_F Fixed)

1. *Increasing product purities* increases reflux ratio.
2. *Increasing feed q* reduces reflux ratio (condenser load) but increases heat input.
3. *Changing feed composition* changes reflux ratio and energy input but not in the same way for all columns. The effect depends on product purities and relative volatilities (see reference 9).
4. *Reducing pressure* usually reduces reflux ratio and energy consumption if product purities are kept constant.

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3

Overhead System Arrangements

3.1 INTRODUCTION

having considered a particular approach to an overall strategy for controlling distillation columns, and having reviewed the fundamentals of distillation, at least for simple columns, let us now turn our attention to some practical aspects. The design of a satisfactory distillation control system involves far more than theory or mathematics. The engineer must have some idea of what constitutes effective equipment configurations and arrangements, as well as an appreciation of equipment performance limitations, and must be able to recognize when undesirable side effects are apt to interfere with an otherwise good control system. Typical equipment, control schemes, problems, and solutions are discussed in this section. The supporting mathematics and theory are covered in Part III.

The column overhead system is generally more complicated than either the feed system or the bottoms system. It usually must condense most of the vapor flow from the top tray, remove inerts, provide reflux flow back to the column, maintain column pressure in the right range, and satisfy part of the column material balance requirements.

Condensate is generally subcooled at least slightly, partly to minimize the likelihood of flashing and cavitation in valves and pumps, partly to control the amount of inerts in the system, and partly to control product losses through the vent. If subcooling is required for a pressure or vacuum column, the preferred method is to have the condensate-temperature controller manipulate the vent flow in some way. This arrangement avoids the instabilities and other control difficulties that often characterize condensate-temperature control systems based on manipulation of the condenser cooling water. In the case of vacuum columns, there may be problems associated with the control of the vacuum jets, and column turndown is usually limited, compared with that of atmospheric or pressurized columns.

Material balance control on the condensate may be accomplished in several ways:

—Flow control of reflux is cascaded, if possible, from column overhead composition, and distillate overflows from the vapor–liquid disengagement space beneath the condenser.

—Same as foregoing except that distillate flow is set by reflux drum level control.

If a smooth flow to the next step in the process is needed, a reflux drum, with averaging level control of distillate, should be employed. If the column top product is a vapor, takeoff should be by “averaging” pressure control. As an alternative, vapor may be taken off on flow control cascaded from top composition control while column pressure is controlled by heat input.

In this book we usually refer to the condensate receiver—the vessel that receives condensate from the condenser—by another name: reflux drum. Although it is less exact, this term is widely used in the petroleum industry.

3.2 TYPES OF CONDENSERS

In chemical and petroleum plants, we find at least five different kinds of condensers:

1. Horizontal shell-and-tube condenser with liquid coolant in the tubes and vapor on the shell side (Figure 3.1). This is probably the most popular type in petroleum refineries. By comparison with the vertical design discussed below, it is much better suited to partially “flooded” operation. In addition, at startup time, column inerts are usually vented more easily (i.e., with less pressure drop) through condensers of this design.

The design illustrated in Figure 3.1 has two vents, each with a valve if the exchanger is operated flooded (see further discussion in Chapter 15). Some designs bring the vapor in at one end and vent uncondensables at the other. Sometimes condensate is taken out through two drawoffs instead of one. The cooling water valve is normally at the exchanger exit to make sure the tubes are filled at all times. Since the exit water is hot, the valve may need anticavitation trim.

2. Vertical shell-and-tube condenser with liquid coolant on the shell side and vapor entering the tubes at the top (Figure 3.2). This type is popular in the chemical industry because it minimizes condenser cost when highly corrosive process materials must be handled.

With a longer condensing path, it is also better suited to applications in which it is desired to absorb the maximum amount of low boilers in the condensate. This condenser commonly has at its lower end a vapor–liquid disengaging pot, which also serves as a condensate receiver. Because all vapors must pass through the tubes, the speed of venting inerts at startup time is limited. For the same reason, this type of condenser cannot be operated partially “flooded.”

Again, the cooling water valve is located at the exit to ensure that the shell is flooded, thereby minimizing corrosion, particularly if 316 SS tubes are used.

3. Internal, in-column head condenser (sometimes called “dephlegmator”). Here we have a number of different designs.

—*Horizontal tube bundle with coolant in tubes* (Figure 3.3). The vapor comes up from below and condensate drops into an annular space around the vapor nozzle. The latter has a “hat” over it to prevent condensate from dropping back down the column. Reflux may return internally via an overflow weir, or externally through a gravity flow line with a control valve.

—*Vertical bundle with coolant on shell side*. This design comes in two variations: “reflux” design, where the vapor goes up the tubes and is countercurrent to the condensate falling down, and a design with a chimney in the center such that the vapor rises in it, reverses direction, and comes down the tubes with the condensate.

—*Air-cooled condensers* (Figure 3.4)

—*Spray condensers* (Figure 3.5). Here condensate is recirculated through

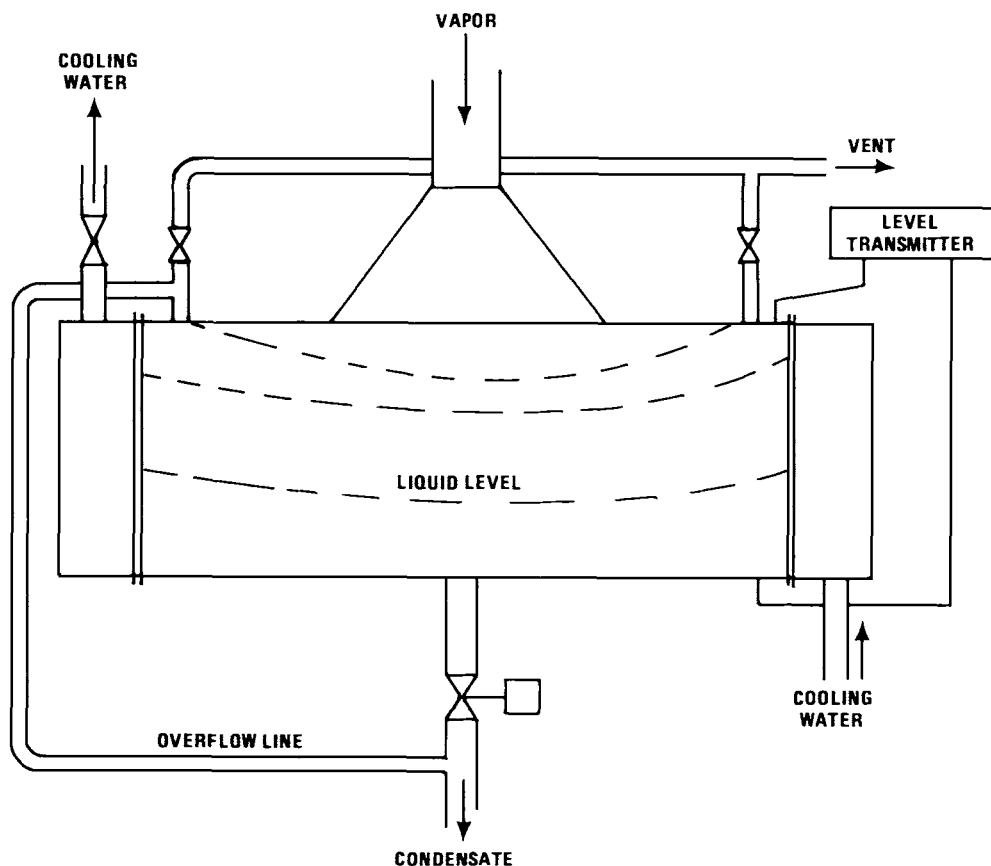


FIGURE 3.1
Horizontal condenser, vapor in shell

a cooler and returned to a spray chamber. This type of condenser is most commonly used in vacuum service because of its low pressure drop.

3.3 ATMOSPHERIC COLUMNS

The preferred overhead system for atmospheric columns is shown in Figure 3.6. The condensed vapor falls into a reflux drum that should have 5–10 minutes' holdup (relative to condensate rate), and inerts are vented to a flare

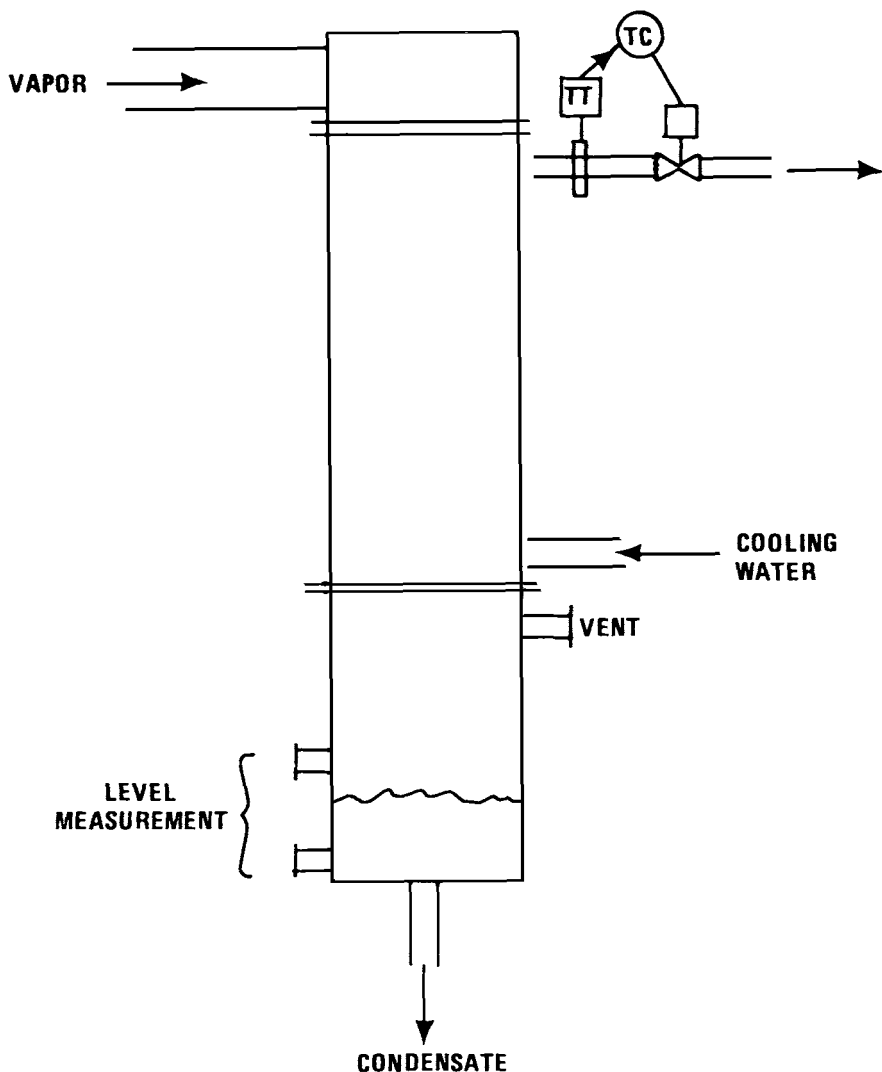


FIGURE 3.2
Vertical condenser, vapor in tubes

or cleanup system. At startup time total reflux may be achieved by using the reflux valve to control the level in the condensate receiver.

For those columns that must be protected from atmospheric oxygen or moisture, a vent system such as that shown in Figure 3.7 should be used. This is similar to the one recommended later for pressurized or vacuum columns. Note that inerts usually should be added *after* the condenser, to minimize product losses. Sometimes, however, it is necessary to add inerts *ahead* of the condenser, for pressure control. Figure 3.7 also shows a more commonly encountered tank arrangement where the reflux drum is common to both the top product system and the reflux system.

A potential and frequent source of trouble with both arrangements is the control of condensate temperature via cooling water. As shown by a study by B. D. Tyreus,¹¹ for a constant subcooled temperature, process gain ($^{\circ}\text{C}/\text{pph CW}$) and dominant time constant both decrease as total heat load increases. This compounds stability problems; we need an increasing controller gain and decreasing reset time as total heat load increases. Further, subcooling heat load must be a reasonable fraction of total heat load—say 5 percent—or the system will lack adequate sensitivity. Finally, many cooling water valves do not have adequate turndown; they are wide open in summer and almost closed in midwinter. A small and a large valve in parallel should often be used.

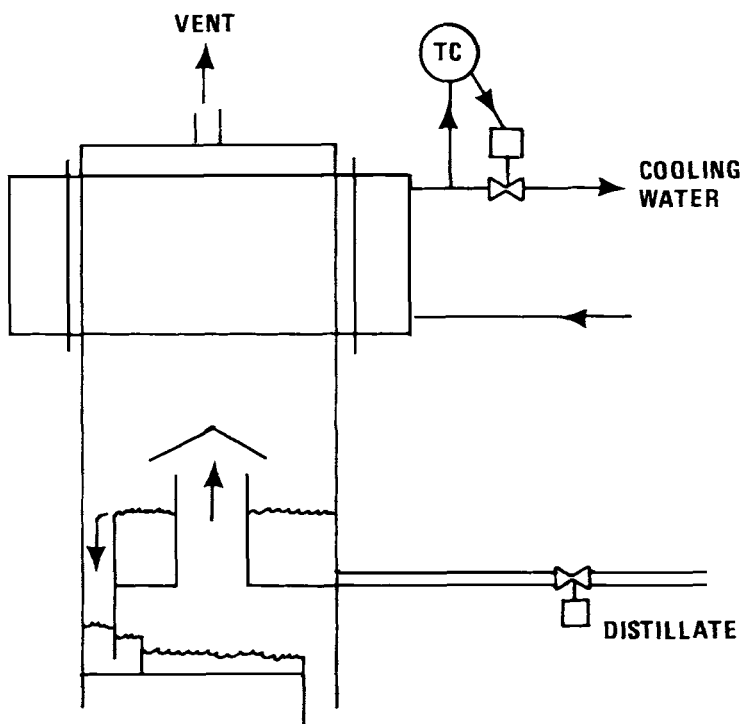


FIGURE 3.3
Alternative overhead system for pressure column

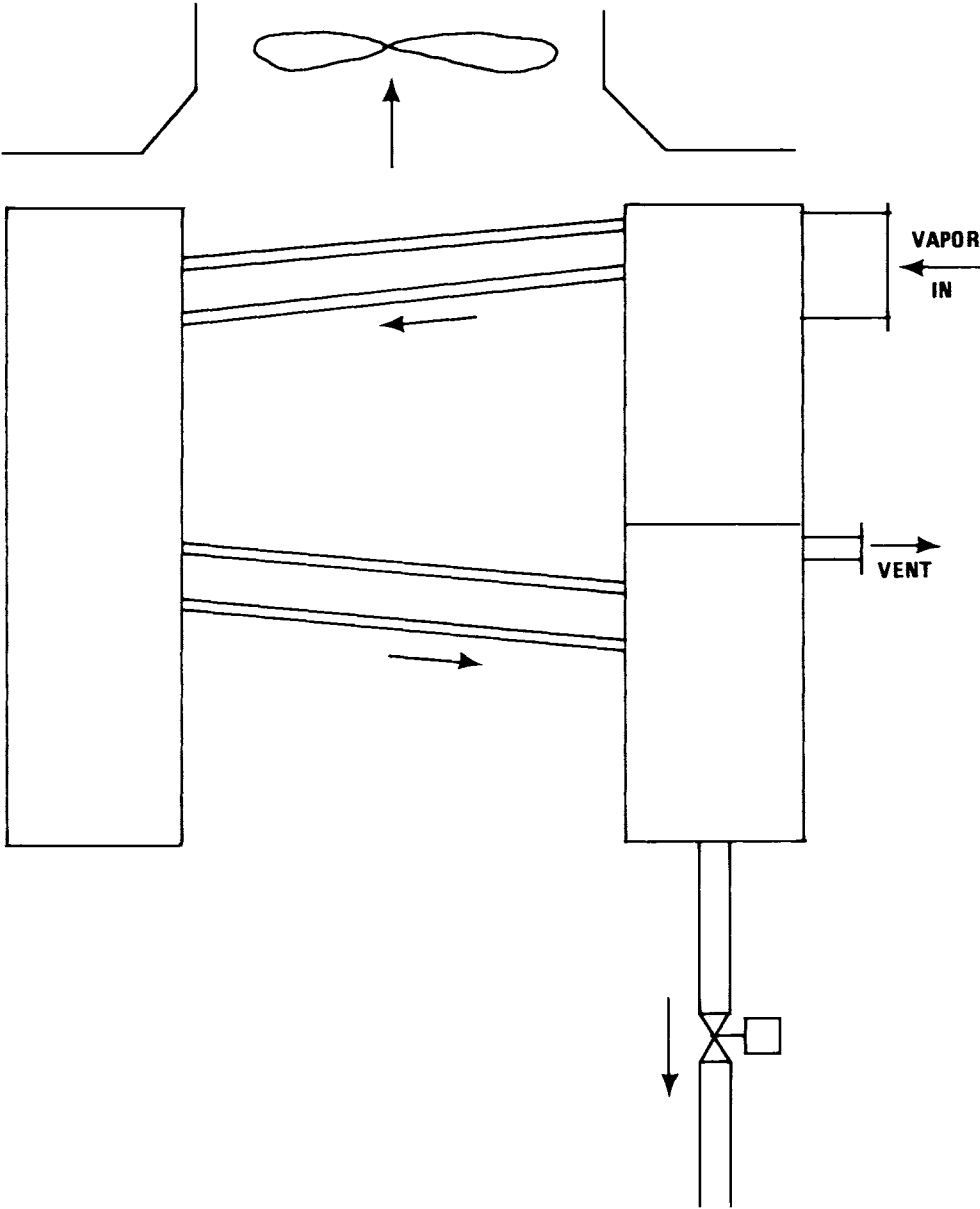


FIGURE 3.4
Air-cooled condenser

The suggested approaches to avoiding these difficulties are as follows:

1. Select the number of degrees of subcooling so that the sensible heat load will be at least 5 percent of the total heat load. This will have a secondary advantage of reducing the probability of cavitation in control valves and pumps.
2. If water-header pressure fluctuations are a problem, use a cascade temperature water-flow control system.
3. If summer–winter heat-load variations are sufficiently severe, use dual, split-range water valves. The smaller valve should open first and will provide adequate winter cooling. The ratio of cooling water rate for the maximum summer heat load to that for the minimum winter load is often two to three times as great as process turndown. (See also discussion in Chapter 11, Section 6.)
4. For the horizontal condenser, the temperature detector preferably should be located in the liquid line just beneath the condenser for maximum speed of response. For the vertical condenser, the temperature detector should be located in a trough at the lower end of a drip collector just below the tube bundle and above the reflux drum. (See Figure 3.8.)

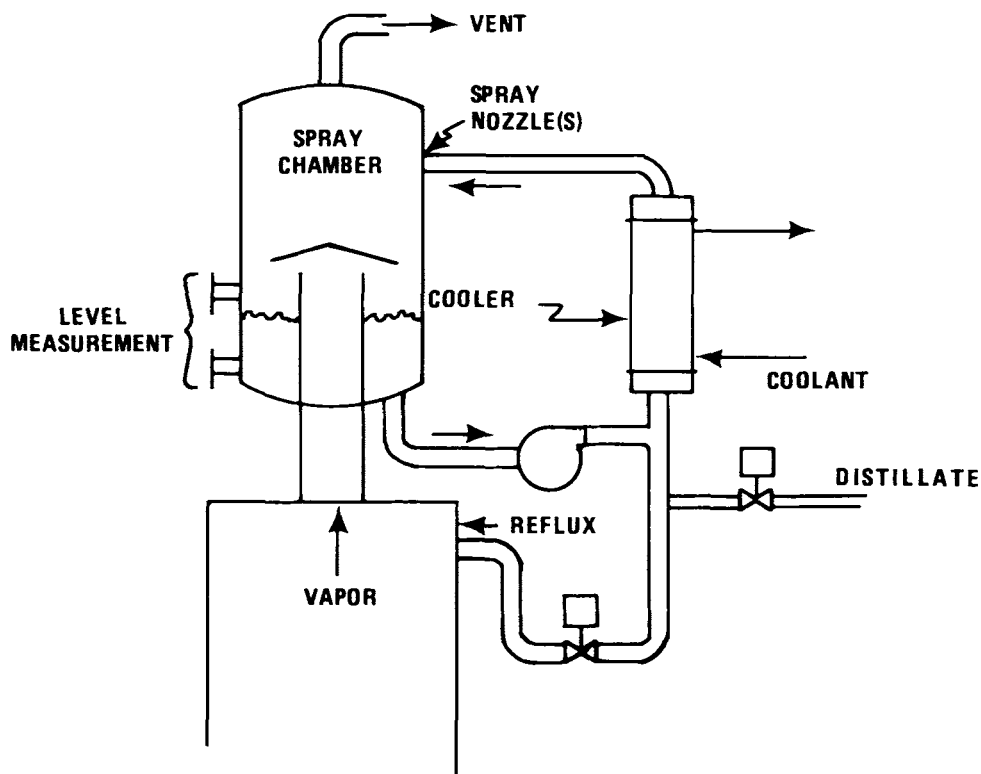


FIGURE 3.5
Spray condenser

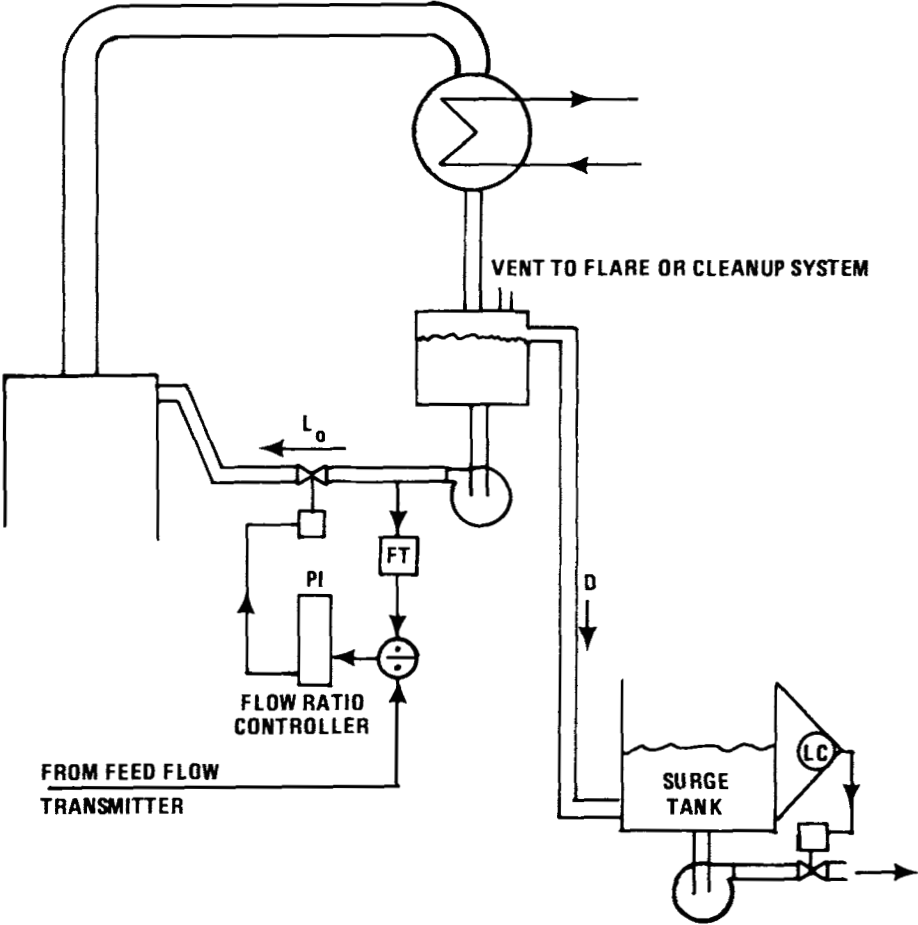


FIGURE 3.6
Preferred overhead system for atmospheric column

5. The controller should have auto overrides (see Chapter 9), or perhaps adaptive gain and reset, to compensate for changes in condenser dynamics as condensate rate changes. An override from cooling-water exit temperature is also normally needed.

6. As an alternative to Item 5, one may use a recirculating coolant system ("tempered" coolant) with condensate temperature control of makeup coolant. This keeps the condenser dynamics constant and eliminates the problem of retuning the controller as the load changes (see Figure 3.9).

7. Another, completely different approach is to run the column at a slight pressure, say 3–5 psig, or under vacuum. Then the condenser cooling water may be manipulated by the pressure controller while subcooling is controlled by manipulating the vent (see Figure 3.10). This is discussed more fully in the next section. There is, however, a limitation to this technique: for protection

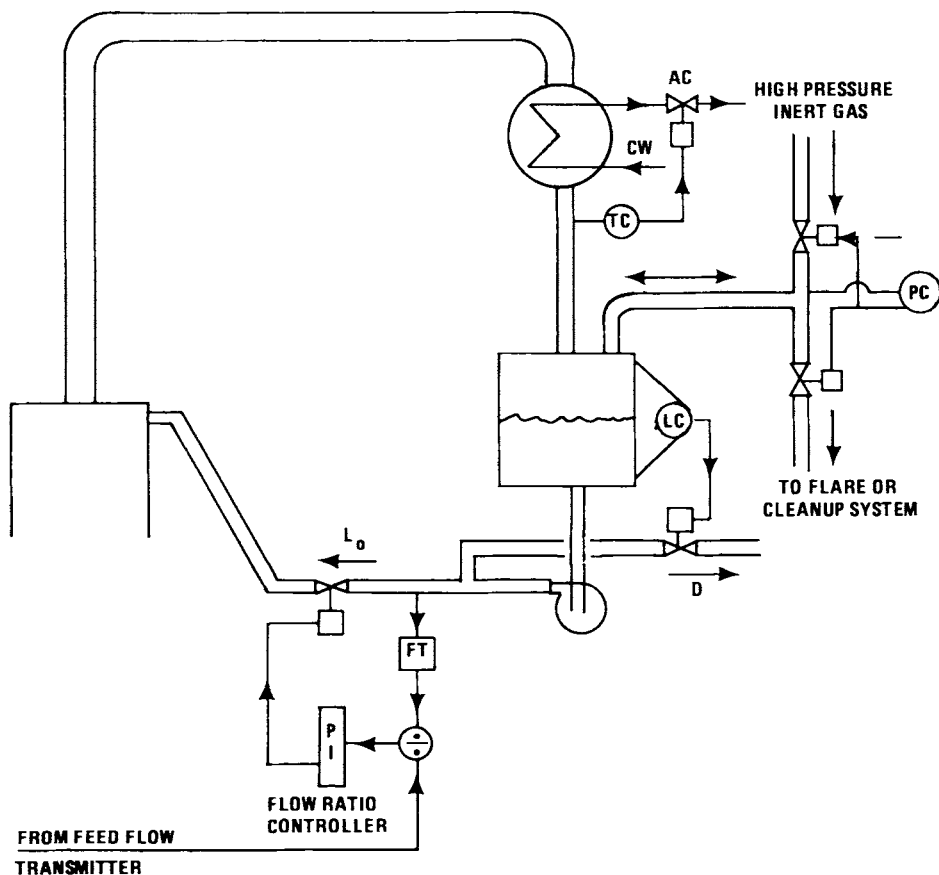


FIGURE 3.7
Alternative overhead system for atmospheric column

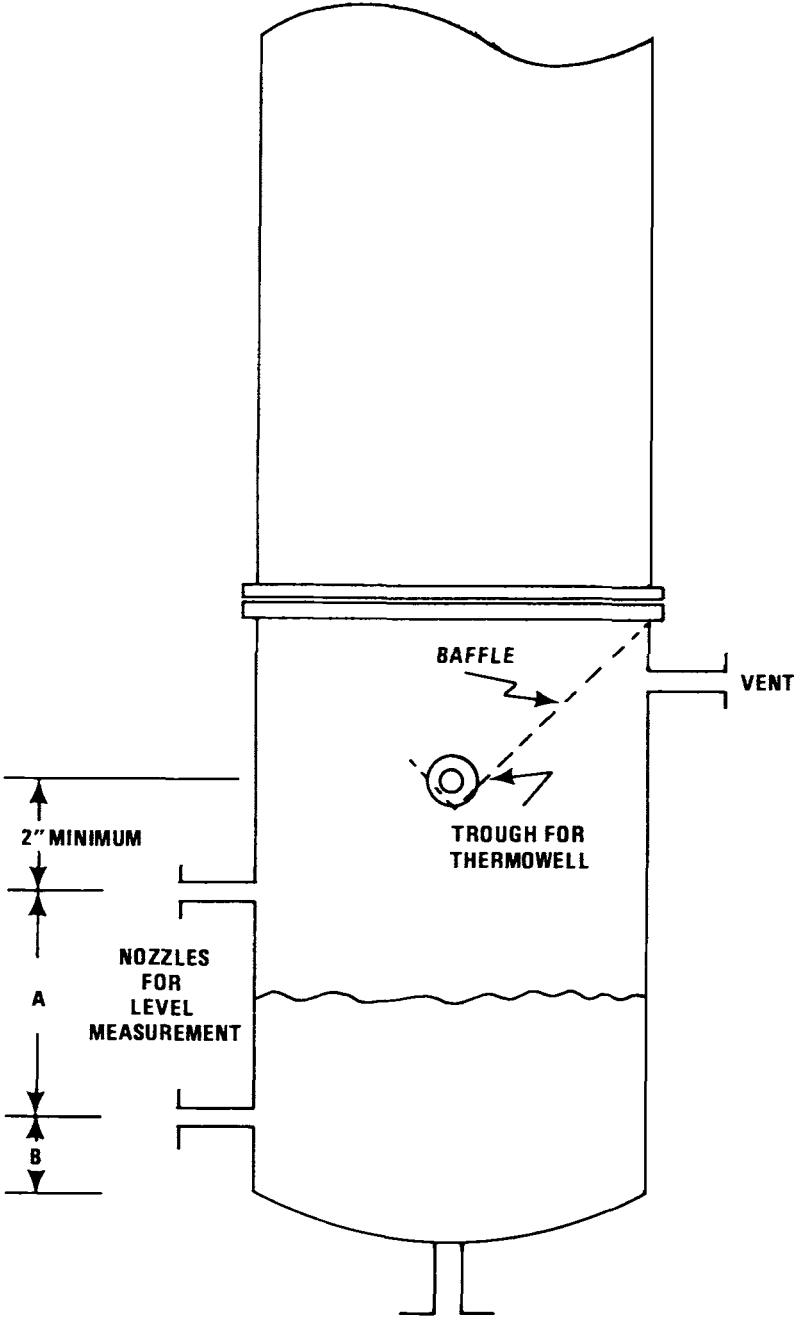


FIGURE 3.8
Thermowell installation under vertical condenser

against fouling and corrosion, cooling water exit temperature should usually be limited to a maximum of 50–60°C (122–140°F), and sometimes lower. Increasingly, override controls are used to provide this protection.

Another alternative is to use exit-cooling-water temperature control, discussed in Section 3.4.

Finally, it is increasingly common to provide no condensate temperature control, but to run with full cooling at all times. This saves a control valve. Further, the quality of cooling water is sometimes so poor that a minimum velocity must be maintained in the exchanger to minimize fouling. For accurate control of internal reflux, an internal reflux computer is required (discussed in Section 11.1). As pointed out by Bolles,¹ however, it is necessary to limit subcooling in some columns to avoid foaming on the top tray.

An additional problem with condensate temperature control, via cooling-water manipulation, relates to column safety. In an instance with which one of the authors is painfully familiar, an atmospheric column with such a control system was running at a very low feed rate. Condensate temperature became too low, so the controller closed the cooling-water valve located in the exit line from a vertical condenser. The water in the shell began to boil, the valve could not pass the required volume of steam, the cooling-water pump stalled, and product vapor issued in great quantities from the vent. Fortunately, an alert operator shut the column down before any damage occurred.

As a consequence, unaided condensate temperature control is not recommended. There should be an override from cooling-water exit temperature. Further, to minimize the hazard of winter freezeup, a limiter should be provided to prevent complete valve closure (see Chapter 9).

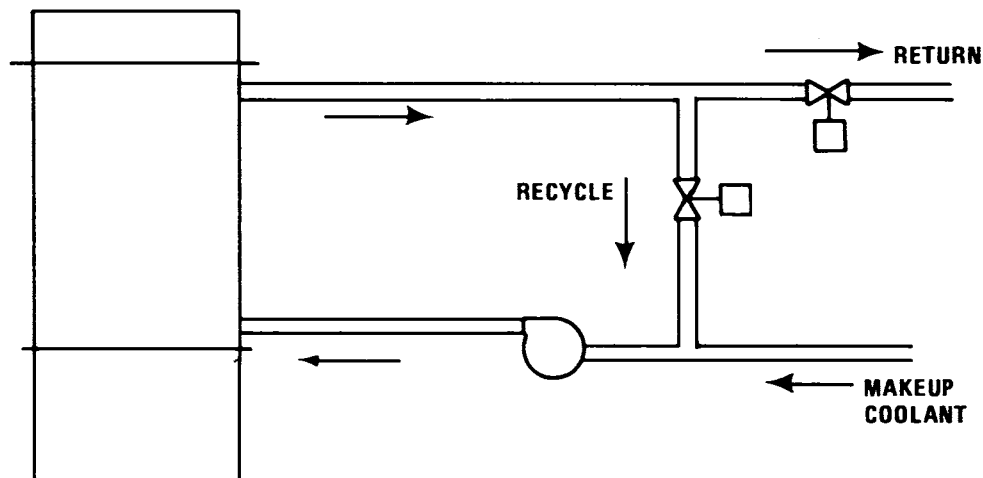


FIGURE 3.9
Tempered coolant system

3.4 VACUUM AND PRESSURE COLUMNS—LIQUID PRODUCT

The preferred arrangement for a vacuum or a pressure column with a large amount of inerts is shown in Figure 3.10. Here the inerts are pulled off or blown out through a vent line in which there is a throttle valve manipulated by the subcooled-condensate temperature controller. For a vacuum column, the low-pressure source is usually a steam jet. If the downstream pressure fluctuates too much, it may be necessary to use a cascade temperature-vent flow-control arrangement.

For a vacuum column with a small amount of inerts, the arrangement of Figure 3.10 may require an impractically small vent valve. In this event the arrangement of Figure 3.11, with a controlled bleed from the atmosphere (or source of inert gas), is better.

A more complicated but more flexible arrangement, such as that of Figure 3.12,* is well suited to either vacuum or pressure columns when the amount of inerts fluctuates over a wide range. It has worked well, for example, on a column in a semicontinuous process that is shut down and started up every day or so, and that must handle severe transients during the startup period. The vent line is connected to a pressure-dividing network with two control valves connected so that as one opens, the other closes. A split-range adjustment of the two positioners (see Chapter 11, Section 10) permits both valves nearly to close when the controller output signal is at its midrange value. Since the sum of the two acoustic resistances is always high, even though each valve is sized to handle a maximum flow equal to five to ten times the average, normal flow of air or gas through the two valves in series is economically small. When an expensive inert gas such as N_2 must be used, it is common to minimize or eliminate split-range overlap to reduce consumption even further. For large columns that must be started up and shut down frequently, an additional large vent valve is sometimes installed in parallel, and split-ranged with the small one. This facilitates getting the column on line at startup.

For many columns the vent flow functions primarily as a purge and is small enough that moderate changes do not affect column operation. In such cases a manually set vent or bleed valve is often adequate and no direct control of condensate subcooling is necessary. For other cases, where column feed rate varies significantly, the vent or bleed valves may be tied to the pressure controller to work in parallel with the condenser cooling-water valve. An example is shown in Figure 3.13.

It should be acknowledged that many engineers today prefer to control condensate temperature by manipulation of condenser cooling water; pressure is then controlled by (1) throttling the vapor takeoff if there is a large amount of inerts, or (2) throttling an air or inert-gas bleed if there is only a small amount of inerts. The objections to, and difficulties with, condensate temperature control, via condenser cooling-water manipulation, were stated earlier. As far

* The symbolism "AO" means air-to-open; "AC" means air-to-close.

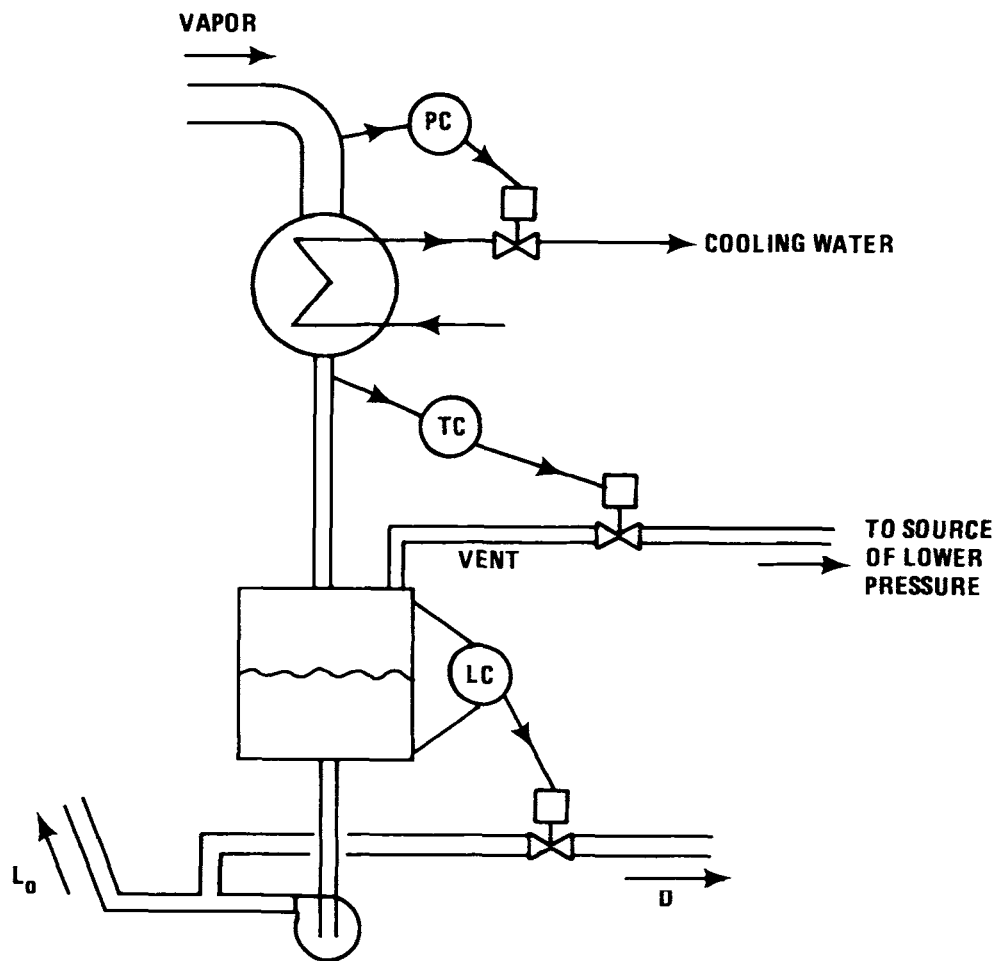


FIGURE 3.10
Overhead system for vacuum or pressure column—large amount of inerts

as we can tell, they are equally valid for pressure and vacuum columns as for atmospheric columns. If cooling water is adjusted manually, the flow is either insufficient or excessive.

So-called “water savers” are cooling-water exit-temperature controls. They have the advantage of minimizing cooling-water flow rate for any given heat load. Their use also minimizes subcooling—and there are instances where this is desirable—but at the expense of variable condensate temperature. This can cause variable internal reflux unless it is compensated for (see Chapter 11, Section 2). This kind of control is often implemented as shown in Figure 3.7; pressure is controlled by manipulation of makeup and vent valves.

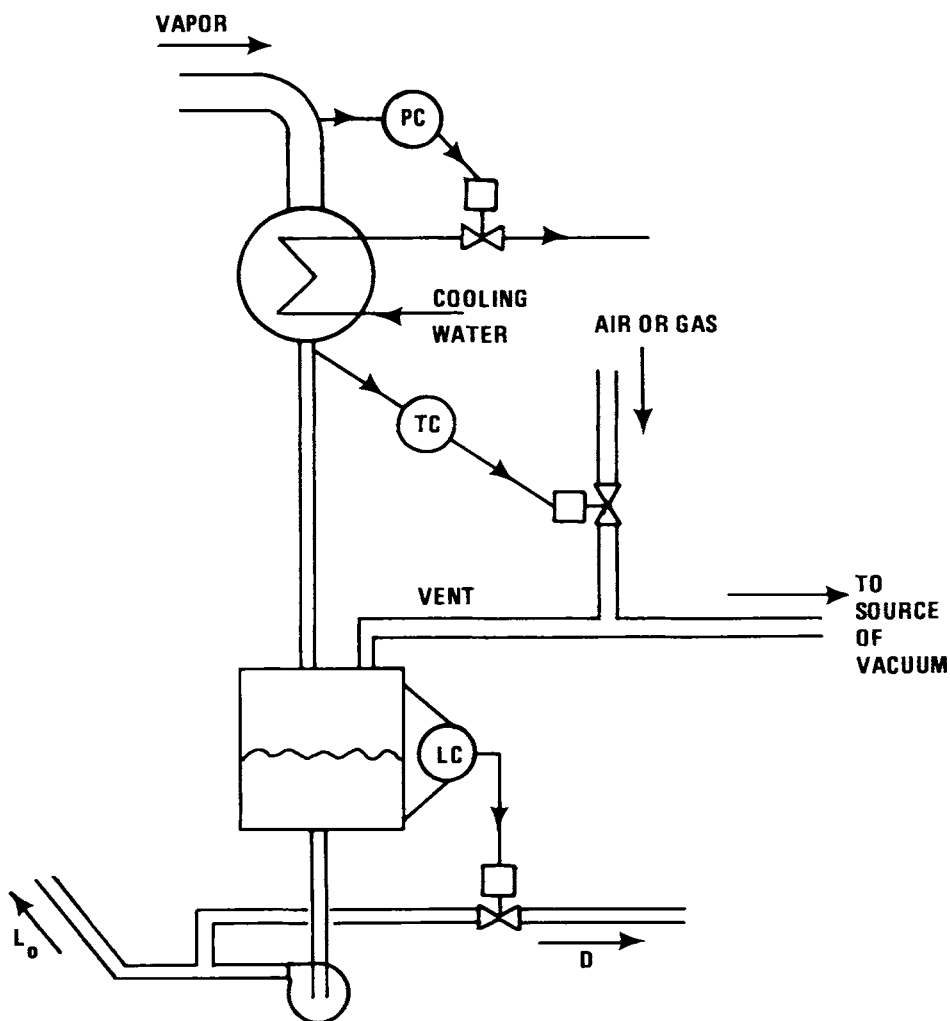


FIGURE 3.11
Overhead system for vacuum column—small amount of inerts

For a large amount of inerts, it is certainly feasible to control pressure by throttling the vent flow; this procedure is recommended for columns with a vapor product (see next section) where condensate temperature is not controlled.

In designing controls for vacuum columns, the engineer should keep in mind that these columns have a narrow range of operation. The range in column pressure drop between flooding and tray instability for a perforated tray column in vacuum service may be no more than 10–15 percent.

One last consideration should be noted here—that of dynamics. We have previously indicated that pressure control, if used, should usually be of the “averaging” type, which provides slow, gradual correction. This fits in well

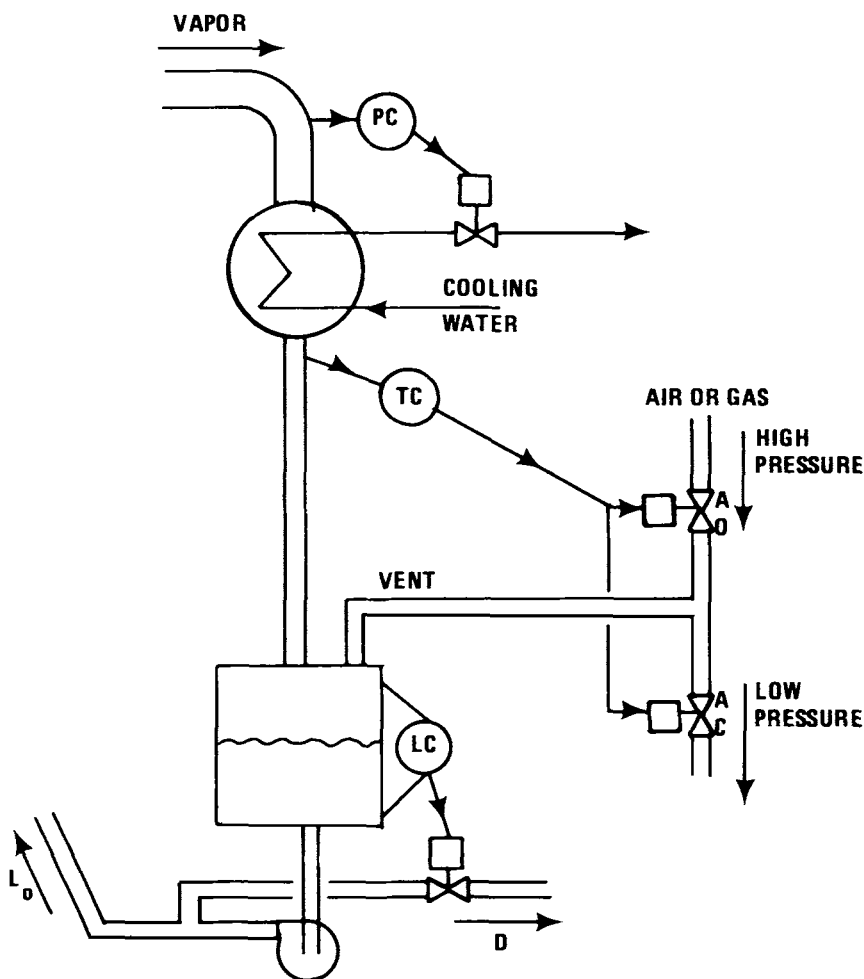


FIGURE 3.12

Alternative overhead system for pressure or vacuum column—small amount of inerts

with condenser cooling-water manipulation since condenser heat loads, like those of most heat exchangers, cannot be changed rapidly. Controlling condensate temperature via bleed manipulation should be comparatively rapid. On the other hand, controlling condensate temperature via cooling-water manipulation requires overcoming the condenser dynamics.

It should be acknowledged, however, that “tight” pressure control is required in some heat-recovery schemes. This is so because process-to-process heat exchangers are often designed for very small temperature differences; small changes in pressure can create relatively large changes in driving force.

3.5 PRESSURE COLUMNS—VAPOR PRODUCT

Pressure columns are sometimes operated so that the product comes off in the vapor phase. If the condenser is external to the column, the arrangement

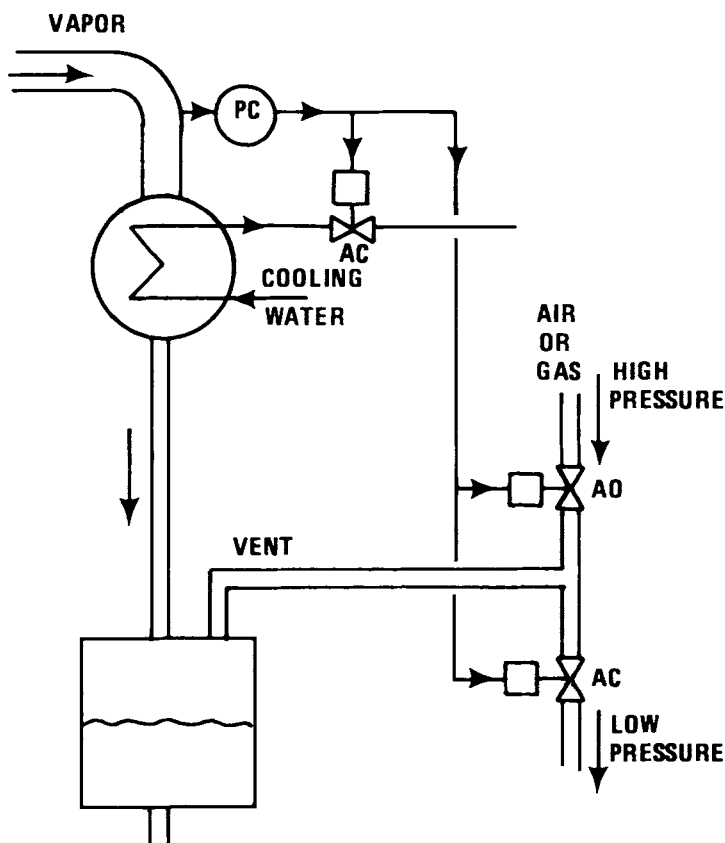


FIGURE 3.13
Alternative pressure control system

of Figure 3.14 may be used. Here column pressure is controlled by manipulating the vapor vent valve. "Averaging" pressure control should be used, and when maximum smoothness of vapor flow is desired, pressure control should be cascaded to vapor flow control. A level controller on the reflux drum balances the rate of condensation against the reflux flow by manipulating condenser cooling water.

An alternative arrangement, used especially when the condenser is built into the head of the column, is that of Figure 3.15. Direct measurement and control of reflux are not possible since the flow is internal. Instead it must be controlled indirectly by manipulation of condenser cooling water, which, in turn, may be reset by a vapor-composition controller. This internal reflux arrangement works well if a heat-computation scheme is used for control. A scheme that we have used successfully is discussed in Chapter 11, Section 4.

If it becomes necessary, because of feed flow or composition fluctuations, or cooling-water-supply fluctuations, to provide reflux-to-feed ratio control, this may be done as follows. By measuring the cooling-water temperature rise

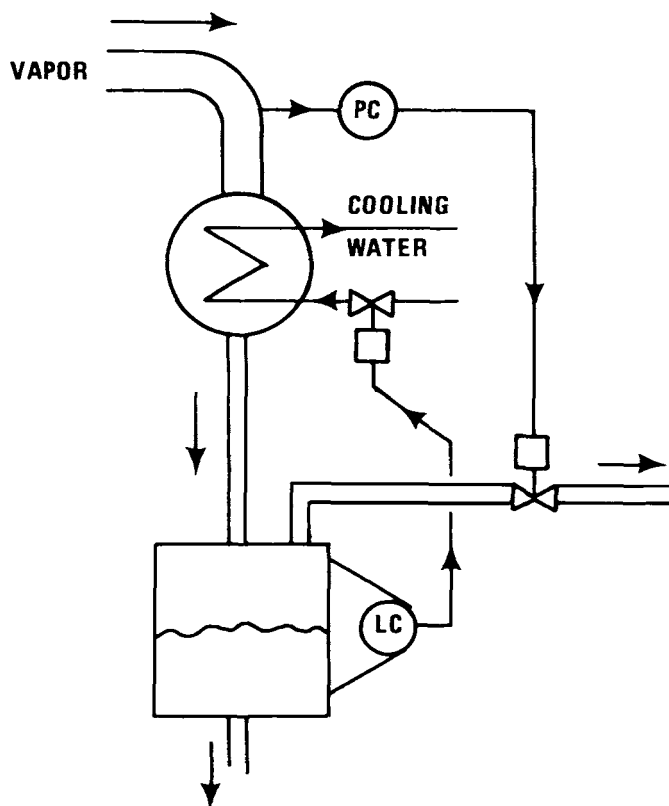


FIGURE 3.14
Overhead system for pressure column—vapor product

and flow rate, we can calculate the heat transferred, q_c . Knowing the latent heat of the reflux, we can calculate w_R , the reflux flow rate in pounds per hour. This calculated w_R can then serve as the measured variable in a reflux flow control system that uses condenser cooling-water flow rate as the manipulated variable.

3.6 MISCELLANEOUS PRESSURE-CONTROL TECHNIQUES

Hot-Vapor Bypass

Another common method for pressure control of pressurized columns involves running with maximum cooling water and bypassing part of the hot gas around the condenser. Several configurations have been employed. In one system the condenser is at a lower level than the receiver by 10 to 15 feet. This means

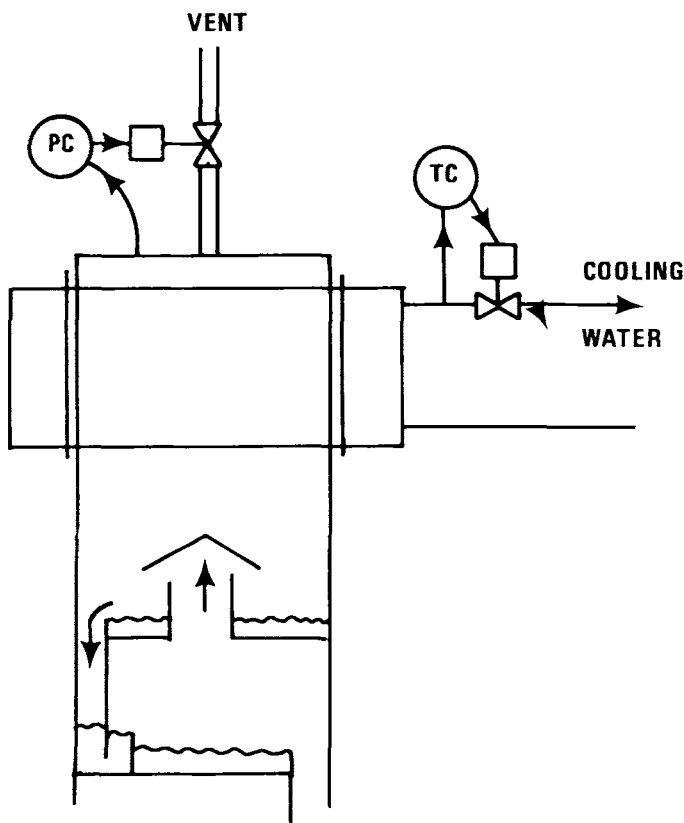


FIGURE 3.15

Alternative overhead system for pressure column—vapor product

that the condenser runs partially flooded. Common practice, as suggested by Hollander² and shown in Figure 3.16A, is to bring the condensate into the bottom of the reflux drum (or at least under the liquid surface) and to bring the hot-gas bypass into the top of the drum. Dynamic problems with such a system can be severe. Suppose, for example, that the column pressure has risen, perhaps as a consequence of increased boilup. The pressure controller pinches the bypass valve to force more vapor into the condenser. This results in a temporary *increase* in pressure since it takes time for the condenser level to drop. Eventually, however, condenser contents drop to a new, lower level, which permits a higher rate of condensation and causes the pressure to be restored. The temporary “wrong-way” pressure response is commonly called “inverse response”; other examples will be discussed later.

Another technique sometimes encountered involves throttling the vapor to the condenser. This suffers from the drawback of requiring a large valve. It also lowers the operating pressure on the condensing side, which limits heat-transfer capabilities.

Another configuration, as shown in Figure 3.16B, has the condenser mounted above the reflux drum. As suggested by Chin³, the hot-gas line around the condenser has *no* valve in it. A valve on the liquid from the condenser floods the condenser to hold column pressure. The liquid in the reflux drum is subcooled, so there is condensation of vapor at the liquid–gas interface in the drum. A vertical reflux drum is recommended to reduce this interfacial area. The liquid line from the condenser should extend down into the liquid in the drum so that the cold liquid is introduced near the bottom of the drum.

These hot-vapor bypass systems are not recommended for systems with even small amounts of inerts.

Flooded Condenser

A pressure-control technique that is growing in popularity involves partial flooding of the condenser without a hot-gas bypass, as shown in Figure 3.17. If the pressure gets too high, the controller opens either the distillate or reflux valve, thereby dropping the liquid level and increasing the heat-transfer area available for condensation. Maximum cooling-water rate is normally used. A mathematical analysis is presented in Chapter 15.

There are some practical problems that must be taken into account. Consider, for example, the horizontal condenser of Figure 3.1. The vapor enters at the center and uncondensed gas exits at the two ends. At low heat-transfer loads, the liquid level will run high. If there is insufficient clearance between liquid level and the top of the shell, violent surging and hammering may ensue.⁴ This may be minimized by designing adequate clearance into the condenser, or by injecting inerts into the incoming vapor partially to blanket the tubes, thereby lowering the liquid level.

Another problem was observed by Mueller⁵ on a partial condenser. At low heat-transfer loads, the liquid inventory in the shell was high. Pressure drop of uncondensed vapor from inlet to the two exits caused a low liquid level in

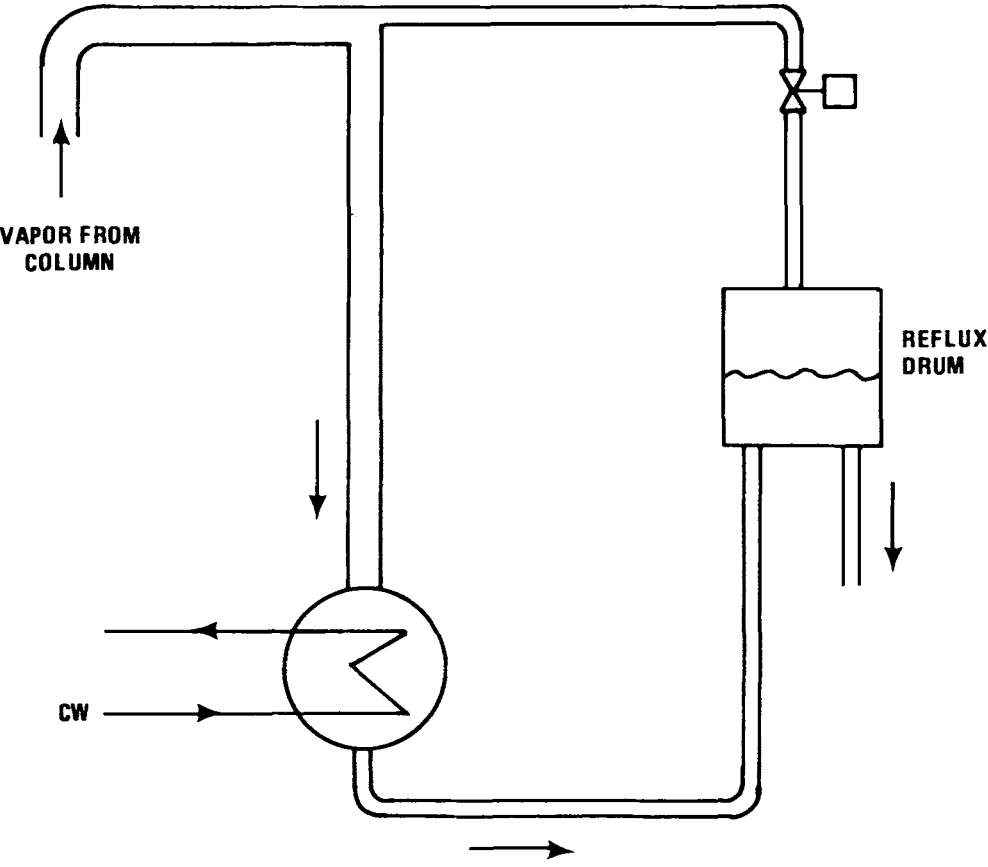


FIGURE 3.16A
Column pressure control by hot gas bypass

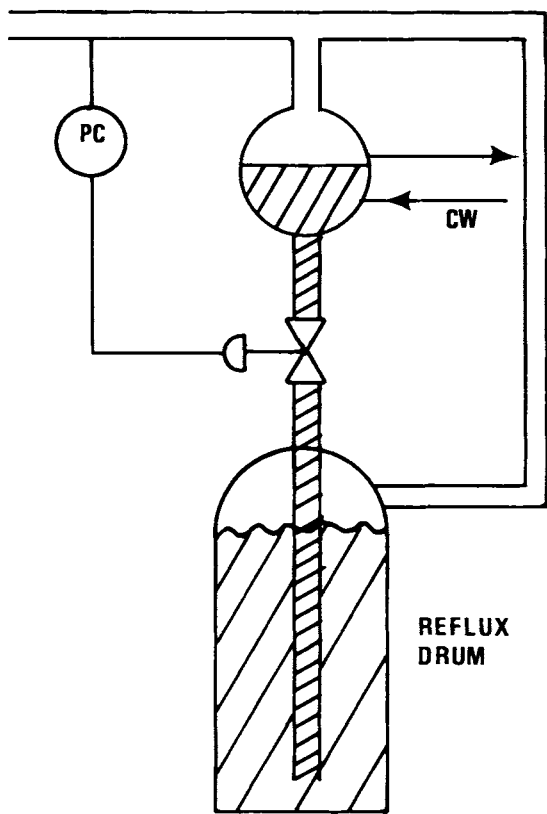


FIGURE 3.16B
Column pressure control by hot gas bypass

the center of the shell and high levels at the two ends. These levels were so close to the exit nozzles that severe entrainment of liquid in the leaving vapor was observed. This particular problem was solved by installing a bypass line between vapor inlet and vapor outlet.

In designing a flooded condenser, one must take care to choose a tube pitch that will not cause large changes in exposed tube area per change in condensate level. In some cases it will be helpful to rotate the tube bundle about its axis just slightly. For troubleshooting, level taps and a level transmitter should be provided. To protect the exchanger from damage at high liquid levels, it will be desirable in some cases to provide an override that will open an inert gas valve connected to the vapor inlet.

3.7 GRAVITY-RETURN REFLUX VERSUS PUMPED-BACK REFLUX

Many arguments have taken place as to whether it is better to locate the condenser overhead and provide gravity return reflux, or to locate the condenser at ground level (or nearly so) and pump the reflux back to the column. With the condenser overhead, the vapor and reflux lines can be short, which favors good overhead composition control. On the other hand, a ground level condenser

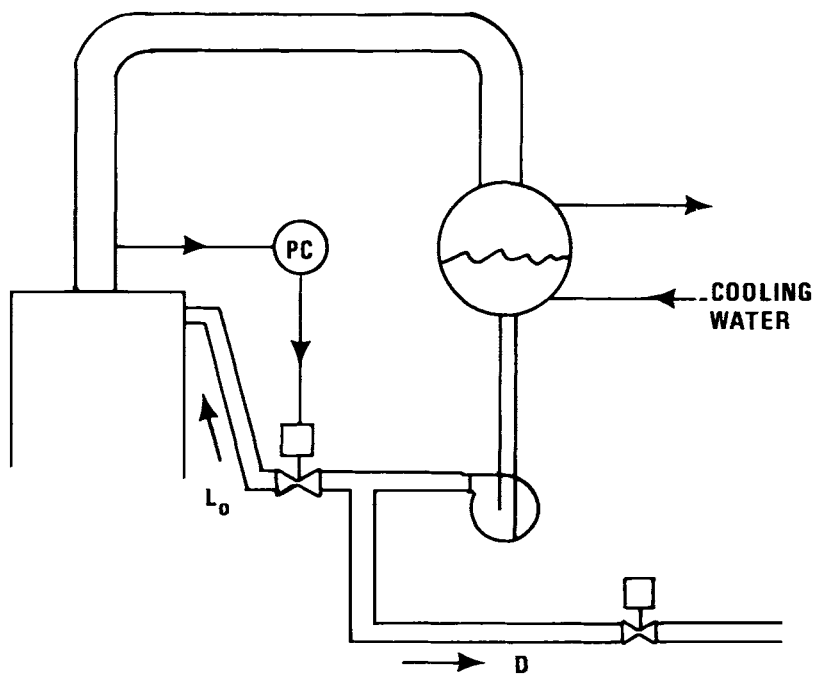


FIGURE 3.17
Column pressure control with flooded condenser

is often easier to maintain, which is especially important if fouling and corrosion are problems. If an overhead condenser is used, a more expensive column-supporting structure is required, particularly if the overhead surge drum is also located at the top of the column. The support problem can be minimized, however, by building the condenser into the top of the column. But one also needs a higher head cooling-water pump, and it is harder to remove the condenser tube bundle for maintenance. If, however, one uses the arrangement of direct return reflux and overflow distillate, then the reflux does not come from the overhead surge drum and this vessel can be located at ground level. Since this tank with its contents is often far heavier than the condenser, the condenser can be located overhead with only a modest increase in structural requirements over a ground-located condenser. Overall a properly designed gravity-flow reflux system is significantly cheaper than a pumped-back reflux system; it is also probably safer since there is no pump to fail. For all gravity-return reflux systems, one must be careful to design the vapor piping and condenser to have a low pressure drop compared with the difference in head between the point of reflux return to the column and the condensate receiver liquid level.

Reflux Flow or Flow-Ratio Control

When reflux is flow or flow-ratio controlled, piping and instrumentation can be very simple. Perhaps the most common arrangement is that of Figure 3.18. Here reflux drum level is controlled by throttling distillate flow. A disadvantage, unless the drum has a large cross section, is that variations in level will cause momentary changes in both reflux and distillate flows. The reflux flow or flow-ratio controller will usually be fast enough that this will not be a problem for reflux flow. The level controller, on the other hand, may have to be cascaded to distillate flow control.

To avoid this problem, one may design a distillate overflow system that provides constant head for reflux. Consider, for example, the scheme of Figure 3.19, which features a vapor-liquid disengagement space built into the lower section of a vertical-tube, coolant-in-shell condenser. For maximum effectiveness the liquid pool in the vapor-liquid disengagement space should have a large cross-sectional area, and the overflow weir should permit a wide range of overflows with only a small change in head. Then, with head across the reflux line fixed, flow will vary only when the valve position is changed. For this application a valve with linear trim will have a linear installed characteristic if line drop is negligible; that is, a plot of reflux flow versus valve stem position will be a straight line. If the individual valve is shop calibrated, then valve stem position can be accurately related to reflux flow.

Use of this technique leads to the arrangement of Figure 3.20. Since the surge tank needs a level controller, there is no savings in instrumentation, but the equipment that needs to be installed at a high elevation is minimized.

Another method of controlling gravity return reflux is shown in Figure 3.21. Here a reflux flow measurement is coupled through a controller to a distillate valve. When this valve is pinched, it backs liquid up into the receiver,

causing more reflux to overflow. An elegant way of doing this is to cause reflux to overflow through a Sutrø weir, as shown in Figure 3.22. The Sutrø weir has the advantage of being a linear weir.⁶

Distillate Flow or Flow-Ratio Control

For those columns with gravity return reflux, a severe oscillation in overhead vapor flow to the condenser is sometimes encountered. This is commonly called “reflux cycle” and has a typical period of several minutes. It has been observed primarily in columns where reflux flow is the difference between rate of condensation and distillate flow rate; that is, where distillate is on automatic flow control or column-composition control. Reversing the controls—that is, employing automatic reflux flow or flow-ratio control and allowing distillate to be the difference flow—provides a positive cure.

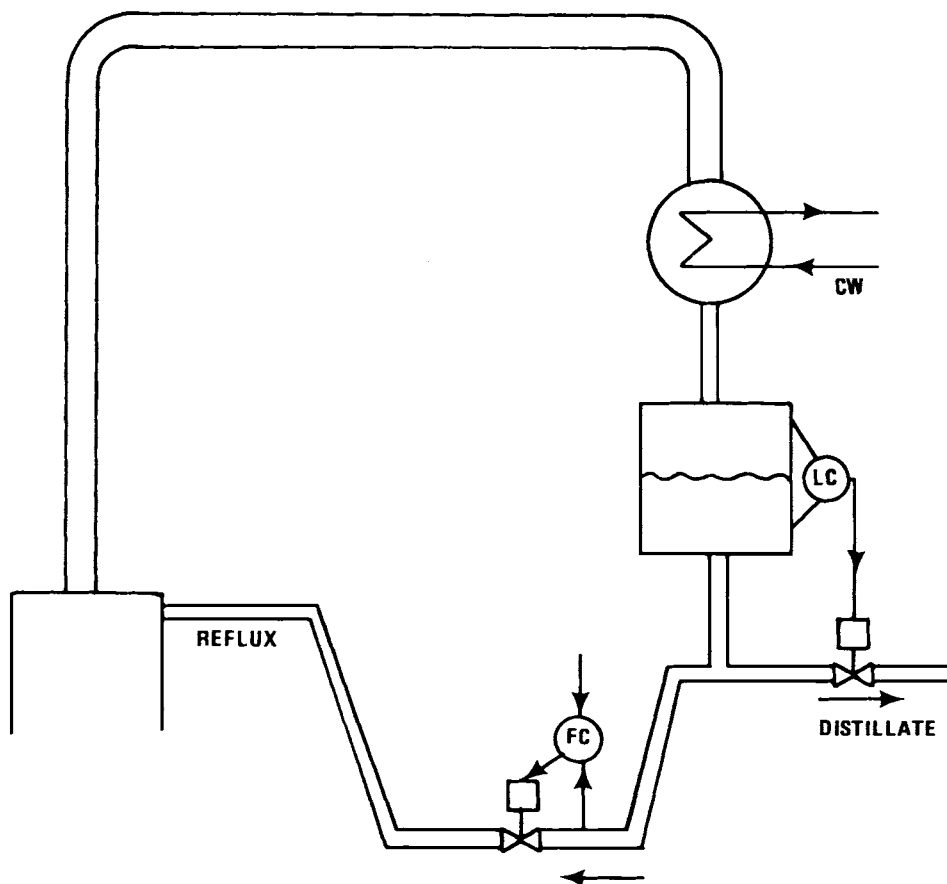


FIGURE 3.18

Gravity flow reflux (flow controlled) and distillate (level controlled)

A mathematical study of this phenomenon has been published.⁷ It was found that the following measures are helpful in increasing stability and minimizing cycle amplitude:

- Provide a difference in head between the liquid level in the reflux drum (or reflux accumulator) and point of reflux return to the column at least ten times as large as the average pressure drop across the vapor piping and condenser.

- Select the flow-metering orifice to hold liquid head in the reflux line at Q_{\max} at the bottom of the pot with the Sutro weir.

- Use a large-diameter vapor line to reduce acoustic resistance and to increase acoustic capacitance.

- Use a horizontal condenser with vapor and a generous free volume on the shell side, or use a short, vertical condenser with many tubes (vapor inside tubes).

- Keep condensate subcooling to a minimum. If subcooling is zero, there will be no reflux cycle.

- Use a condenser with a recirculating coolant.

- Increase column operating pressure. This increases vapor density and decreases $\partial T_c / \partial P$, thereby further improving stability; T_c is the condensing temperature.

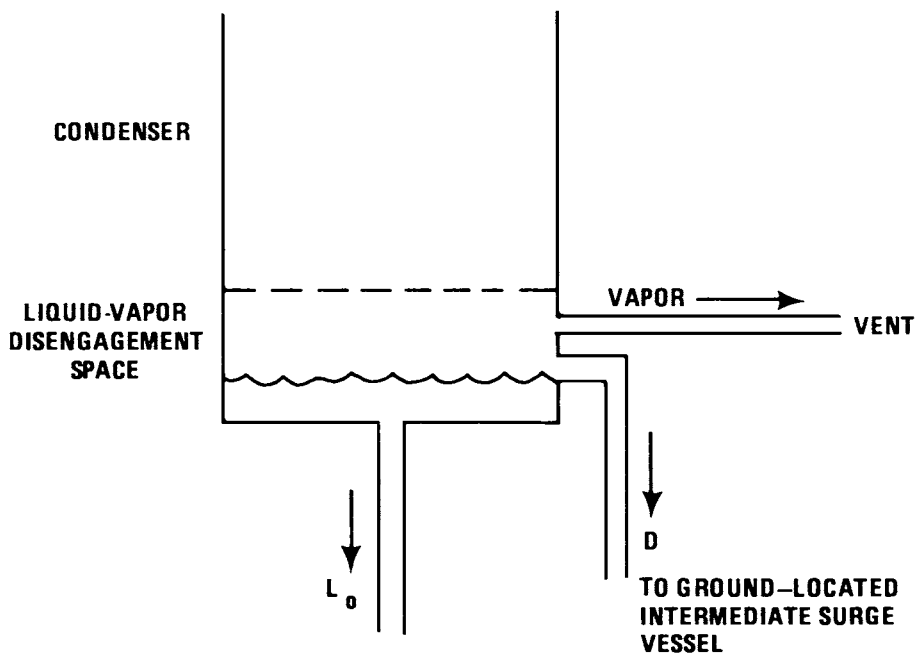


FIGURE 3.19

Liquid-vapor disengagement space built into condenser

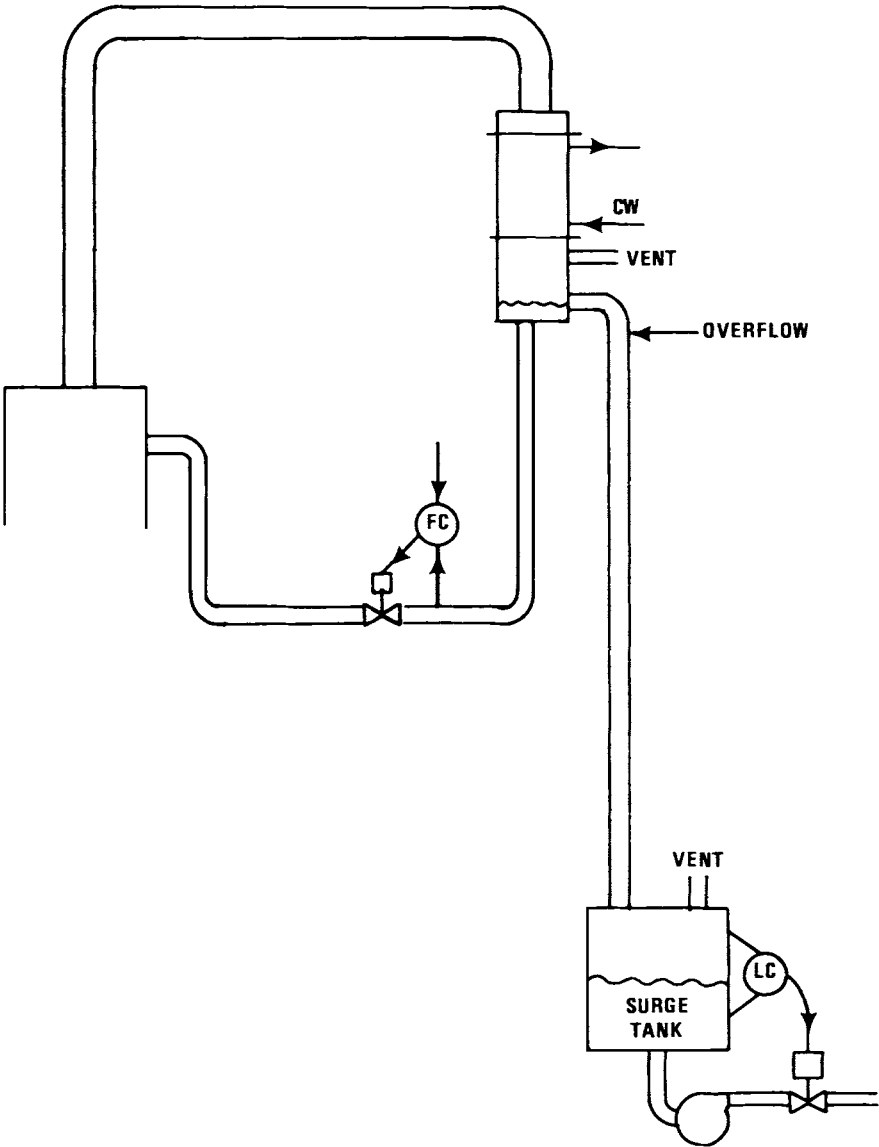


FIGURE 3.20
Gravity-flow reflux system with ground-located surge tank for distillate

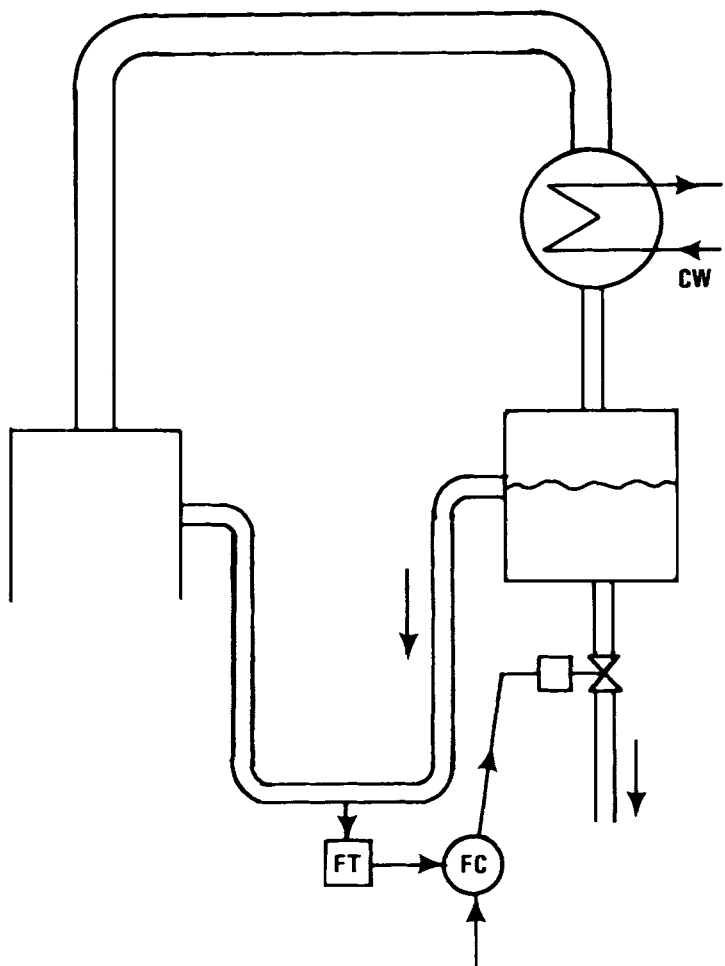


FIGURE 3.21
Control of gravity reflux flow rate by throttling top product flow

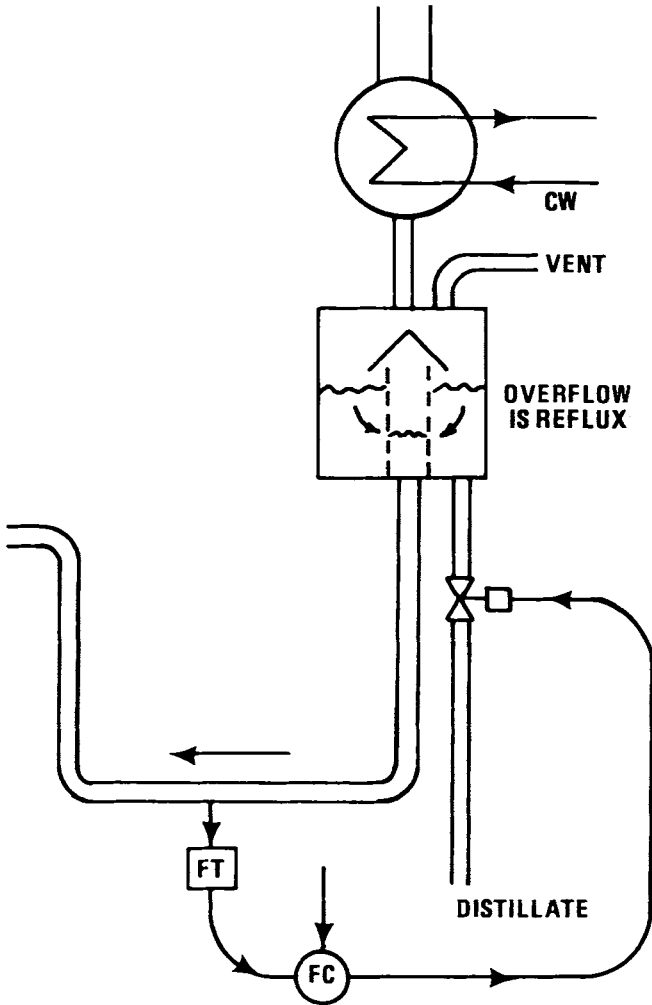


FIGURE 3.22
Control of gravity reflux flow rate by overflowing through Sutrö weir and by throttling distillate flow

Although the preceding are helpful, it is sometimes necessary to take stronger measures. The following list of overhead schemes is in order of preference, from best to least desirable, when reflux flow must be the difference flow between rate of condensation and distillate flow:

—Pumped-back reflux, proportional level control cascaded to reflux flow control, $\tau_H \geq 3-5$ minutes. τ_H is the level control time constant (see Section 3.10).

—Pumped-back reflux, proportional level control, $\Delta p_v \geq 10$ psi, $\tau_H \geq 3-5$ minutes.

—Gravity-flow reflux, proportional level control cascaded to reflux flow control, $\tau_H \geq 3-5$ minutes.

—Gravity-flow reflux, proportional level control, $\Delta p_v \geq 5$ psi, $\tau_H > 3-5$ minutes.

—Gravity-flow reflux, surge tank with Sutro weir, $\tau_T > 3-5$ minutes. Make sure the reflux line has a sufficiently high hydraulic resistance. Note that τ_T is the hydraulic time constant of the surge tank with Sutro weir:

$$\frac{Q_o(s)}{Q_i(s)} = \frac{1}{\frac{A}{\partial Q / \partial H} s + 1} \quad (3.1)$$

$$= \frac{1}{\tau_T s + 1} \quad (3.2)$$

where

$\frac{\partial Q}{\partial H}$ = constant for a Sutro weir

Q_o = outflow, ft³/min. from Sutro weir

Q_i = inflow from condenser, ft³/min.

A = cross-sectional area of vessel, ft² (vertical, cylindrical design assumed).

The last scheme is shown in Figure 3.23. Plant experience indicates that it does not completely eliminate the cycling, but reduces the amplitude by a factor of ten or more to an acceptable value. It is simple and inexpensive to fabricate and permits locating the condenser at a lower elevation than do any of the other techniques.

Another important point for gravity return reflux is the method of connecting the reflux piping to the column. Each of the two piping arrangements of Figure 3.24 has an undesirable upward loop just before entry into the column. Inerts sometimes accumulate in this pocket, thereby causing a reflux flow oscillation as a result of an intermittent siphon action. There have been cases where hot vapor was sucked back into this pocket and caused such severe hammer that the reflux line and column nozzle were ruptured.

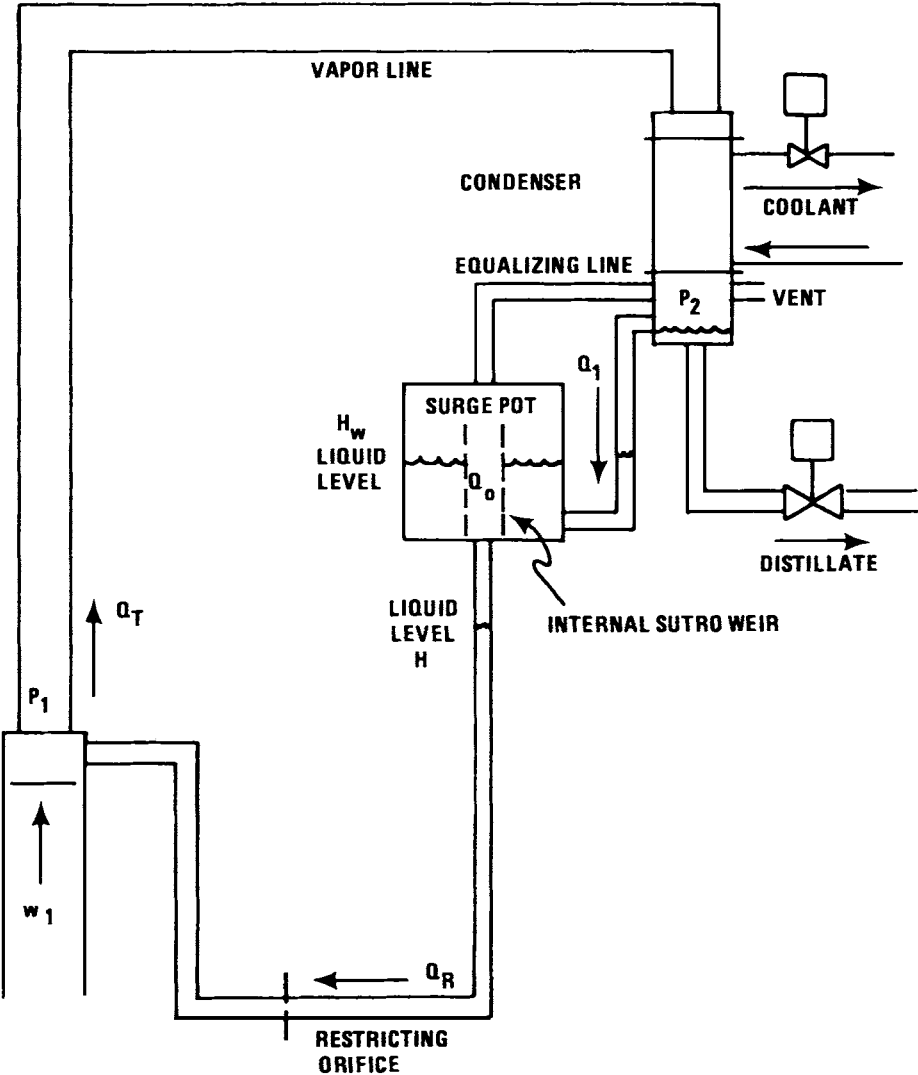


FIGURE 3.23
Gravity-flow reflux, surge tank with Sutro weir, $\tau_T > 3\text{--}5$ minutes

This phenomenon is particularly troublesome with vacuum towers where some slight air leaks are unavoidable. The preferred arrangement of Figure 3.25 avoids this kind of flow instability; the piping may enter horizontally, or with a slight inclination as shown.

3.8 CONTROL TECHNIQUES WITH AIR-COOLED CONDENSERS

In recent years air-cooled heat exchangers have grown enormously popular. They have demonstrated, however, certain control problems. They are far more

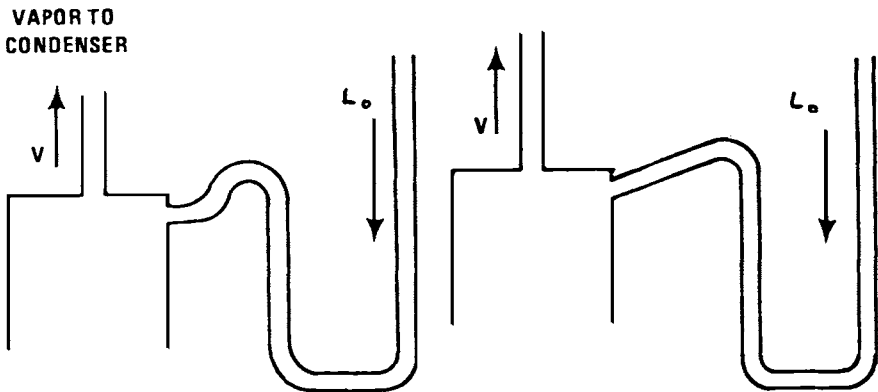


FIGURE 3.24
Undesirable piping arrangements for returning reflux to column

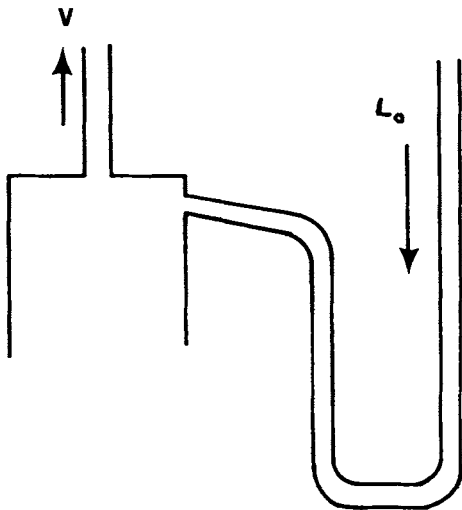


FIGURE 3.25
Preferred piping arrangement for returning reflux to column

sensitive to atmospheric changes such as rainstorms or even changes in wind velocity than are liquid-cooled exchangers. Various techniques have been devised to control the rate of heat transfer or to compensate for condensate temperature changes:

1. Use of induced-draft rather than forced-draft exchanger designs. Top-located fans provide much better protection against rainstorms (see Figure 3.4).
2. Partial bypass of hot liquid from upper section of the exchanger and mixture with cold liquid leaving at the bottom. This permits sensitive, rapid temperature control.
3. Variable-pitch fans.
4. Adjustable louvers in the exchanger housing to control suction air flow.
5. Internal reflux computers (see Chapter 11).
6. Flooded operation.
7. Variable-speed fan drives.

3.9 “TEMPERED” VERSUS ONCE-THROUGH COOLANT

The term “*tempered*” has been applied to coolant systems that feature a high circulation rate through the condenser as shown by Figure 3.9. The temperature rise per pass is kept small, and the condenser must be designed for a small pressure drop on the coolant side to minimize pump horsepower requirements. A high-flow, low-head pump therefore is required. The two control valves may be replaced by a single three-way valve if the recirculating flow is not too much larger than the return flow.

Tempered coolant is employed for either or both of two reasons:

1. It eliminates problems with high-freezing-point condensate that might plug the condenser if once-through coolant were used. In extreme cases tempered coolant is taken from and returned to a supply tank that is temperature controlled.
2. Condenser dynamics are radically improved over those achieved with once-through coolant. Speed of response is greater and condenser dynamics do not change with load changes. Condensate-temperature and column-pressure control are easier.

3.10 LEVEL CONTROL OF CONDENSATE RECEIVER AND REQUIRED HOLDUP

In this section, comments or suggestions regarding required holdup will be primarily from the standpoint of getting good, or at least adequate, control of the column with which the holdups are associated. Small holdups favor good composition control. But when the holdups are part of a feed system for another process step, requirements may be much greater. This is discussed in more detail in Chapter 5.

Level control in condensate receivers or reflux drums is commonly achieved by manipulating either top product flow or reflux flow. Less commonly, overhead level control is accomplished by adjusting boilup or by adjusting condenser cooling water. For the first two cases, a relatively simple control system can be used.

For maximum flow smoothing, it uses the cascade PI level-control to flow-control scheme of Figure 3.26. For this example level is maintained by throttling distillate flow. Note that the PI level controller must be enhanced with high—and low—overrides (called “auto overrides”) to keep level within the vessel. (With electronic analog or microprocessor controls, an alternate design with nonlinear gain and reset may be used—see reference 12.) Since, however, there are two outflows, one must also have overrides on reflux for the same reason. The quantitative design is discussed in Chapter 16. Note that the flow measurement must be linear (or linearized) for stability reasons. Cascade control is used to eliminate flow changes caused by control-valve upstream and downstream pressure variations.

For level control via reflux-flow manipulation, it is necessary to sacrifice flow smoothing in the interest of good composition control. If a PI controller

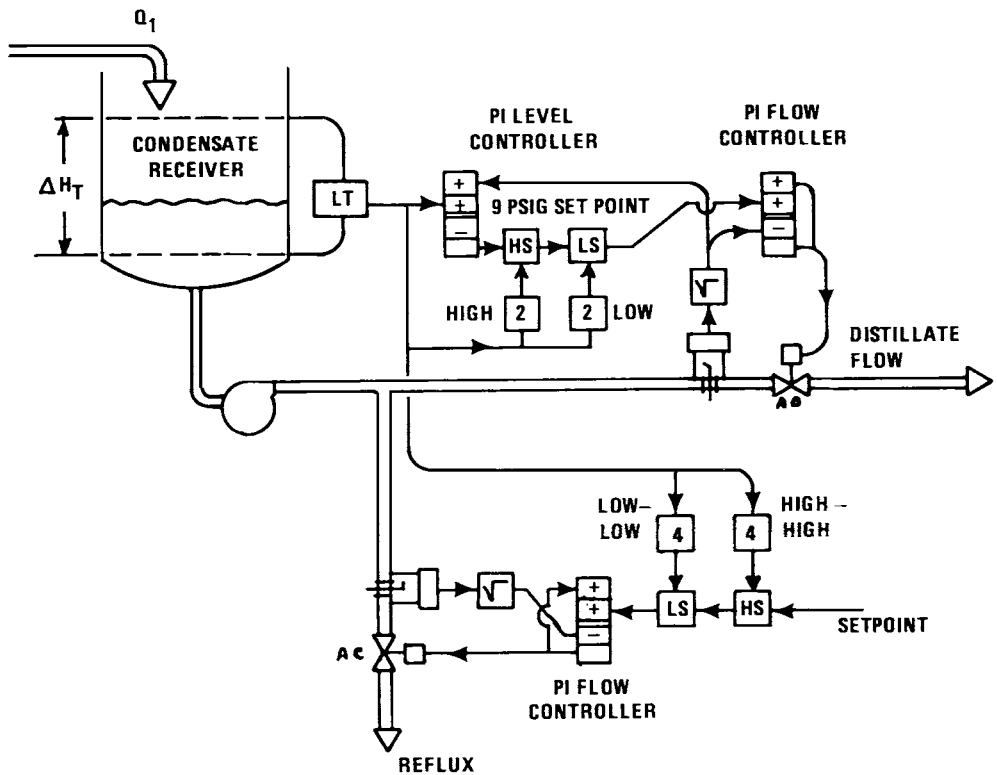


FIGURE 3.26
Condensate receiver level control via distillate

is used (usually without overrides), it should be tuned for tight control of level, not averaging level control. For this application it is probably more appropriate to use a proportional-only controller as shown in Figure 3.27.

As indicated, it uses a controller with gain 2 (50 percent PB). For pneumatics the bias is so set that the output is 9.0 psig when the input is 9.0 psig. This means that the control valve is closed at the 25 percent level and wide open at the 75 percent level. These numbers should be regarded as part of process design, and the bias adjustment therefore should be treated as a calibration adjustment rather than as a "tuning" adjustment. For pneumatic systems inexpensive fixed-gain relays are available for this application. Figure 3.27 also shows simple overrides that act on the distillate valve. If level gets too high, the distillate valve is opened; if level becomes too low, the distillate valve is closed. Chapter 9 discusses overrides further.

If the manipulated valve has a linear installed flow characteristic (preferred), and if there is no level self-regulation (if Δp_v does not change appreciably with change in level),* then the dynamic response of the proportional-only† level control system may be defined by a first-order time constant:

$$\tau_H = \frac{A}{K_{mb} K_c \frac{dQ_o}{d\theta_c}} \quad (3.3)$$

where

τ_H is in minutes

A = cross-sectional area, ft^2 , of seal pot (vertical, cylindrical design assumed)

K_{mb} = level transmitter gain

$= \frac{\Delta\theta_{mb}}{\Delta H_T}$; ΔH_T is the level transmitter span corresponding to the output signal span $\Delta\theta_{mb}$ (psi for pneumatics)

K_c = controller gain, dimensionless

$\frac{dQ_o}{d\theta_c}$ = valve gain, $\frac{\text{ft}^3/\text{min}}{\text{psi}}$

$= k_v(Q_{FS}/\Delta\theta_c)$ for valve with linear installed flow characteristics; $\Delta\theta_c$ is input span of valve positioner corresponding to full valve travel

Q_{FS} = flow-sheet value of manipulated flow, ft^3/min

* If there is significant level self-regulation, one should use cascade level-flow control.

† As will be seen in Chapter 16, τ_H is also important in the design of PI level controls. It is expressed a little differently, however, for cascade controls.

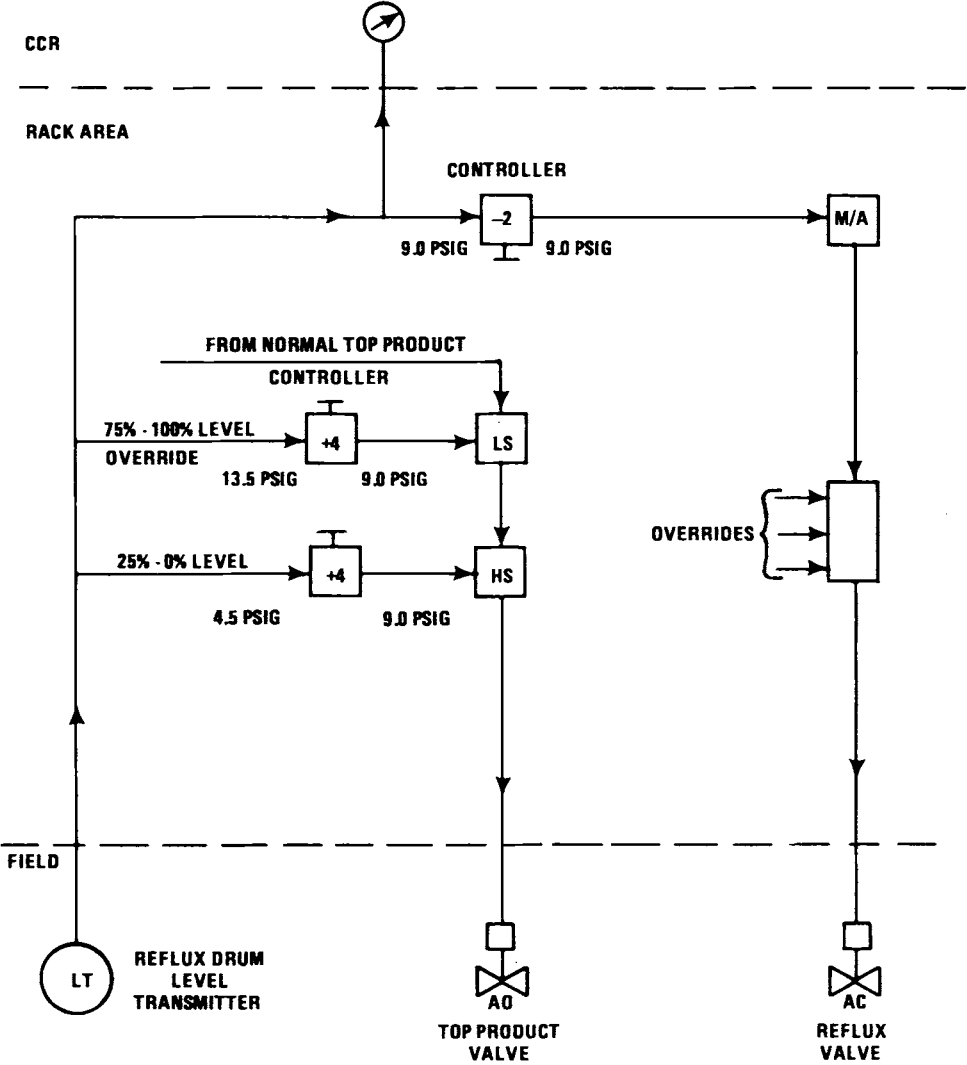


FIGURE 3.27
Proportional-only condenser seal pot level control via reflux flow

During the early stages of a design project, the level nozzle spacing usually must be determined before control valves are sized. Typically, however, valve-sizing procedures lead to:

$$(Q_o)_{\max} = k_v Q_{FS} \quad (3.4)$$

where k_v is a multiplying factor typically in the range of 2–6. (See discussion on valve sizing in Chapter 11.)

Then, from equation (3.3), for a linear installed valve characteristic, for $k_v = 4$, $K_c = 2$, and $\Delta\theta_{mh} = \Delta\theta_c$:

$$\Delta H_T = \frac{K_c k_v Q_{FS}}{A} \tau_H \quad (3.5)$$

$$= \frac{2 \times 4 Q_{FS}}{A} \tau_H = \frac{8 Q_{FS}}{A} \tau_H \quad (3.6)$$

In the discussion that follows, the various control schemes usually will require a $\tau_H \geq 2$ minutes. For any given scheme, however, one should make sure that override time constants are at least 1 minute, that is, $[\tau_H]_{OR} \geq 1$ minute. This may require a τ_H much greater than 2 minutes.

If pneumatic instruments are involved, the preceding is adequate for two-pipe designs with up to a 1200-foot one-way distance for 1/4-inch OD plastic tubing or a 2000-foot one-way distance for 3/8-inch OD plastic tubing. For electronic-analog or microprocessor controls, the limiting factor will be the speed of response of the valve-positioner valve-actuator combination.

If, for process reasons, the available holdup must be very small, it is sometimes possible, by the use of special techniques, to design for τ_H or $[\tau_H]_{OR}$ less than 1 minute. High-performance valve positioners probably will be required, and for pneumatic instruments various methods are available for minimizing lags and improving speed of response.⁹ Experimental data for long pneumatic tubing runs are given in reference 10. System performance then should be checked by frequency-response methods or computer simulation. As noted earlier, if additional volume is required for smoothing out feed to the next process step, it should be preferably in a separate vessel outside of the reflux path.

Process engineers often think of “holdup time” rather than a time constant, τ . Holdup time is usually considered to be equal to volume divided by throughput. If we think in terms of the volume corresponding to the level transmitter span,

$$\text{Holdup time} = \frac{A \Delta H_T}{Q_{FS}} = \tau_{HU} = \tau_H \times K_c \times k_v$$

$$\tau_H = \frac{A \Delta H_T}{\Delta\theta_m \times K_c \times \frac{k_v Q_{FS}}{\Delta\theta_c}}$$

and

$$\Delta\theta_m = \Delta\theta_c, \quad K_c = 2, \quad k_v = 4$$

$$\tau_H \cong \frac{A \Delta H_T}{8 Q_{FS}}$$

Therefore, if $\tau_H \geq 2$ minutes,

$$\begin{aligned} \tau_{HU} &\geq 2 \text{ minutes} \times 2 \times 4 \\ &\geq 16 \text{ minutes} \end{aligned}$$

For proportional-reset controllers, we will usually use $K_c = 0.25$. Then:

$$\begin{aligned} \tau_{HU} &\geq 2 \text{ minutes} \times 0.25 \times 4 \\ &\geq 2 \text{ minutes} \end{aligned}$$

This illustrates an advantage of PI controllers in averaging level-control service. For the same τ_H , only one eighth the volume is required.

There is some disagreement about optimum reflux drum holdup. Small holdups of liquid are desirable from the standpoint of reducing time constants in the overhead composition control loop. This permits faster and tighter composition control.

Larger reflux drum holdups (10–30 minutes in terms of total condensate rate) are favored by some designers because they provide more liquid-surge capacity. This enables the column to ride through larger disturbances without losing reflux flow, and consequently internal liquid flows (which take some time to reestablish). In the experience of one of the authors, larger reflux drum holdups have proved particularly desirable for columns that occasionally experience slugs of light ends or inerts in the feed. The condenser is essentially blanketed during the period it takes to vent these noncondensibles off. Without several minutes of reflux holdup, liquid flows would be lost and the time for the column to recover from this upset would be appreciably lengthened. This problem, however, may be minimized by appropriate use of overrides.

Level Control Via Top Product (Distillate)

If level goes high, indicating that the sum of the distillate and reflux flows is less than condensate rate, we want to increase the reflux flow. As shown in Figure 3.26, this is accomplished by a relay with a gain of 4 and a high selector (HS). On the other hand, if level goes too low, we want to pinch reflux flow. This is accomplished by another gain 4 relay and a low selector (LS). (See Chapter 9 for further discussion.) Let us now see what effect the difference between top product flow and reflux flow has on τ_H and $[\tau_H]_{OR}$. Let us suppose that $Q_R = 5 Q_D$, where Q_R is reflux flow and Q_D is top-product flow. Then:

$$\tau_H = \frac{A \Delta H_T \times 12}{K_c \times 12 Q_{mfd}} \quad (3.7)$$

and

$$[\tau_H]_{OR} = \frac{A \Delta H_T \times 12}{K_{sc} K_{OR} \times 12 Q_{mFR}} \quad (3.8)$$

where

Q_{mFD} = distillate flow-meter span, ft³/min

Q_{mFR} = reflux flow-meter span, ft³/min

and K_{OR} is the override relay gain and K_{sc} is the subcooling factor discussed in the next section. Here it is assumed that both the distillate- and reflux-flow control loops are fast compared with the level control loop. If $K_c = 0.25$, $K_{sc} = 2$, and $K_{OR} = 4$:

$$\tau_H = \frac{A \Delta H_T \times 12}{0.25 Q_{mFD} \times 12} \quad (3.9)$$

and

$$[\tau_H]_{OR} = \frac{A \Delta H_T \times 12}{4 \times 2 \times Q_{mFR} \times 12} \quad (3.10)$$

Therefore,

$$\frac{\tau_H}{[\tau_H]_{OR}} = 32 \frac{Q_{mFR}}{Q_{mFD}} \quad (3.11)$$

If the flow-meter spans are in the same ratio as the average flows (5:1), then:

$$\frac{[\tau_H]}{[\tau_H]_{OR}} = 32 \times 5 = 160 \quad (3.12)$$

This indicates that override action may be extremely rapid compared with that of normal level control. This is not usually desirable; it may upset the process. For this reason we mostly choose $K_{OR} = 2$ if possible.

Level Control Via Reflux Flow

For the system of Figure 3.27, this is very similar to the previous case except that the controller must have a gain of -2 for an AC reflux valve and a level above 75 percent must open the distillate valve, while a level below 25 percent must close it. For gravity flow reflux, level control via reflux should be avoided, if possible, since such designs are often plagued by a "reflux cycle" as mentioned in Section 3.7. If this design cannot be avoided, the entire condenser-reflux system should be designed according to the recommendation in Section 3.7.

For level control via reflux, the characteristic time constant is defined a little differently:

$$\tau_H = \frac{A \Delta H_T}{K_{sc} K_c (Q_R)_{FS} k_v} \quad (3.13)$$

where

$$K_{sc} = \text{subcooling constant, } \frac{\text{lbm internal reflux flow}}{\text{lbm external reflux flow.}}$$

$$= 1 + \frac{c_p}{\lambda} (T_O - T_R) \quad (3.14)$$

where

c_p = reflux specific heat, pcu/lbm °C

λ = vapor latent heat, pcu/lbm

T_O = vapor temperature, °C

T_R = external reflux temperature, °C

To allow for condenser dynamics, τ_H should be at least 5 minutes and $[\tau_H]_{OR}$ at least 2 minutes. Note that reflux valves should be sized to handle the maximum rate for total reflux operation.

For columns with simple controls, level control via reflux has the advantage that external reflux temperature changes do not change internal reflux.

Level Control with Small Seal Pot Volume

If satisfactory flow smoothing cannot be achieved with a gain 2, proportional-only level controller (usually because available holdup is very small), one should use a proportional reset level controller. The proper design is discussed in Chapter 16.

Level Control Via Boilup

For overhead level control via boilup, a dynamic analysis should be made to determine proper holdup and controller type. If level is cascaded to flow control, the flow transmitter should have a linear output with flow. If an orifice ΔP transmitter is used, this should be followed by a square root extractor.

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4

Column-Base and Reboiler Arrangements

4.1 INTRODUCTION

the design of a column base with its associated reboiler can be a complex matter. It requires simultaneous consideration of fluid mechanics, heat transfer, mass transfer, and process control. For example, the following must be considered when a vertical-thermosyphon, forced-circulation, or kettle-type reboiler is used:

1. Spacing between the vapor-return nozzle and the lowest tray should be large enough to minimize entrainment of liquid drops in the rising vapor. This is typically one vapor nozzle diameter and normally will be specified by the column designer.

2. The maximum liquid level should not be too close to the vapor nozzle as this will promote turbulence in the liquid surface and liquid entrainment into the rising vapor.

3. Minimum liquid level should not be too far below the vapor nozzle, or the falling liquid drops will entrain too much vapor into the liquid pool. In severe cases this may cause foaming in the column base and “gassing” of the reboiler. This difficulty, however, can be largely offset by item 4.

4. There should be a minimum liquid level depth above the nozzle to the drawoff line from the column base or vaporizer separator. This is required for two reasons: first, to permit entrained vapor bubbles to rise and separate from the liquid pool, and second, to minimize the likelihood of vortex formation at the drawoff nozzle. Vortexing (like the swirl at a bathtub or sink drain) is undesirable since it promotes entrainment of vapor into the drawoff line, which may cause “gassing” of the reboiler or bottom product pump or circulating pump. Vortex breakers should be installed routinely.¹

5. If level-measurement nozzles are not protected by an internal damping chamber, they should have an orientation no more than 90° from the vapor-return nozzle and should not be located under the last downcomer.

In this chapter we consider primarily five types of reboilers: vertical thermosyphon, flooded thermosyphon (steam side), forced circulation, flooded-bundle kettle, and internal reboilers.

4.2 VERTICAL THERMOSYPHON REBOILERS*

Typical Operation and Design

Vertical-thermosyphon reboilers or calandrias are commonly designed for a mass circulation ratio of about 5–20 parts of liquid to one part of vapor. This means that only about 5–25 percent by weight of reboiler effluent is vapor; on a volumetric basis, however, the numbers are almost reversed.† The tubes are filled with froth, and pure liquid exists in the tubes only for a few inches above the bottom tube sheet.

For most applications, if shell-side steam pressure, tube-side pressure, and composition are fixed, heat-transfer flux will be a maximum when column-base liquid level relative to the bottom tube sheet is about one third of the distance between the two tube sheets²—see Figure 4.1A.

If liquid level goes below this point, heat transfer falls off rapidly. As liquid level goes above this point, heat transfer declines only slightly; at a liquid elevation corresponding to the top tube sheet, flux is perhaps 10–15 percent less than maximum. In view of this, it is a conservative practice to maintain column-base liquid level below the top tube sheet and above an elevation corresponding to the midpoint between the two tube sheets. As pointed out by Smith,³ the above may be influenced by column pressure.

For many applications the practical implementation of the preceding is as illustrated by Figure 4.1B. A level nozzle spacing of 44 inches is suggested for ΔP transmitters and 48 inches for displacer transmitters. The top nozzle center should be at least 1 inch below the lip of the vapor-return nozzle, or 1 inch below the elevation of the bottom lip of the internal vapor downspout (if used). The bottom nozzle should be centered 6 inches above the KRL (knuckle radius line) if a ΔP transmitter is used, and 2 inches above the KRL if a displacer instrument is used. For either type of transmitter, the following calibration procedure is recommended.

—Zero level should be 6 inches above the KRL.

—The transmitter output span should be 3–15 psig for pneumatics, or 4–20 mA for electronics, for $44 - 6 = 38$ inches of process fluid. This allows 6 inches below the top level nozzle to accommodate changes in liquid-specific gravity.

* Particularly in the petroleum industry, horizontal-thermosyphon reboilers are sometimes encountered.

† A literature review of thermosyphon reboilers is given by McKee.⁴

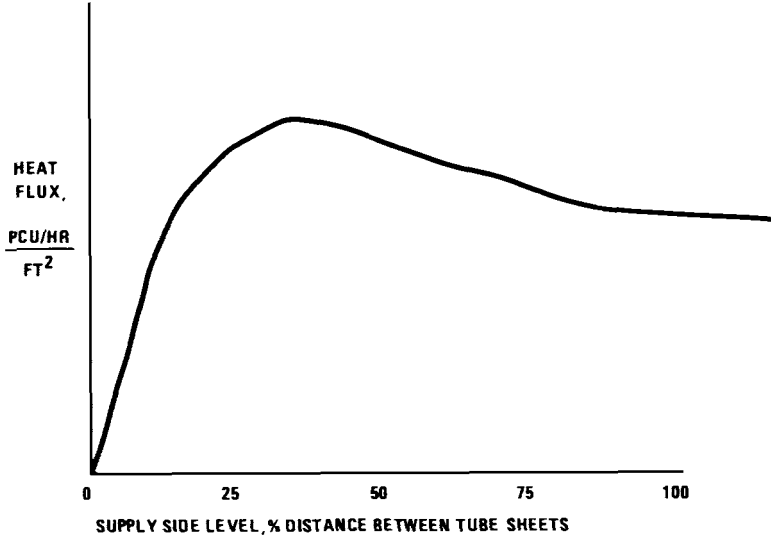


FIGURE 4.1A
Vertical thermosyphon—heat flux vs. supply side liquid level

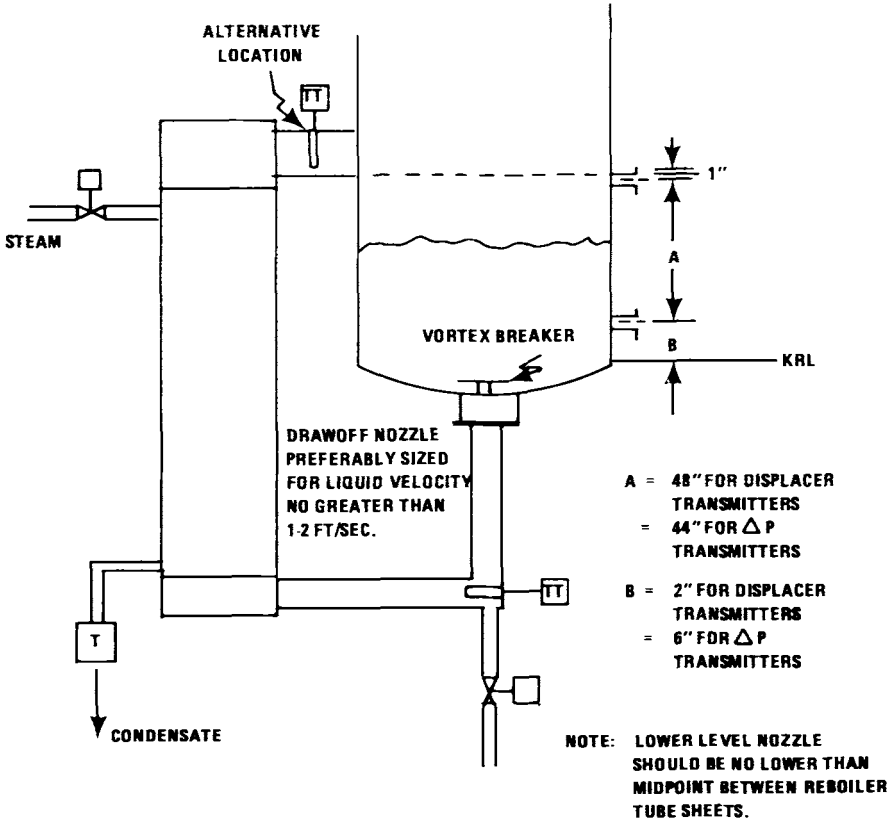


FIGURE 4.1B
Distillation column base with thermosyphon reboiler

Since we normally design the controls to work in the middle of the level transmitter span, the average level will be $6 + 19 = 25$ inches above the KRL.

Level-measurement techniques, including baffle design, are discussed in Chapter 11.

Control and Operation Difficulties

A typical thermosyphon reboiler has several additional features that sometimes cause or contribute to control difficulties:

A. Choked-Flow Instability

When the reboiler heat flux

$$\frac{\text{pcu/hr}}{\text{ft}^2}$$

or

$$\frac{\text{Btu/hr}}{\text{ft}^2}$$

becomes too high, there ensues a violent oscillation of process flow through the reboiler tubes and of reboiler ΔP . In one set of experiments,⁵ a period of 3–4 seconds was observed; it is not known how general this is. Corrective measures include reducing the vapor exit restrictions and increasing the liquid supply restriction.

B. Reboiler–Column-Base “Manometer”

From a hydraulics standpoint, a reboiler and column base function crudely as a U-tube manometer whose natural period in seconds is approximately $\sqrt{l}/1.3$ where l is the length in feet of the liquid column. Manometer theory says that damping is increased as the liquid flow restriction is increased. There may be a connection between this and choked-flow instability; for a period of 3 seconds, $l = 15$ feet, which is fairly typical for industrial installations.

C. Change in Tube Vapor Volume with Change in Steam Rate

For a fixed liquid level in the column base, the volumetric percent vapor in the tubes increases as steam (or other heating medium) flow rate is increased. As shown by Figure 4.2, this effect is very pronounced at low heat loads but tapers off at higher loads. The end result is similar to boiler swell (Chapter 10, reference 6); an abrupt increase in steam flow causes liquid to be displaced back into the column base, with a resulting temporary increase in column base level.

Figure 4.2 is derived by running the reboiler computer design program for at least three values of base level and five values of heat load. The program calculates the weight fraction vapor at the end of each of a number of sections of tube length. The cubic feet of vapor in each section of tube are then calculated from the following equation:

$$[\bar{V}_v]_i = \frac{1}{1 + \frac{\rho_v}{\rho_L} \left(\frac{1 - \bar{F}_{wt}}{\bar{F}_{wt}} \right)} \times L \times V'_i$$

where

ρ_v = vapor density, lbm/ft³

ρ_L = liquid density, lbm/ft³

$$\bar{F}_{wt} = \frac{[F_{wt}]_{i-1} + [F_{wt}]_i}{2}$$

$[F_{wt}]_i$ = weight fraction vapor at exit of tube section i

L = length, in feet, of tube section

V'_i = inside volume of one tube section, ft³/ft

The total vapor volume in the tubes is then $\Sigma[V_v]_i \times n$ where n is the number of tubes and $\Sigma[V_v]_i$ = total vapor volume, ft³, in one tube.

D. Change in Tube Vapor Volume With Change in Column-Base Liquid Level

If heat load is held constant but column-base liquid level is varied, tube vapor volume decreases with increase in liquid level. In cases examined so far, this effect is small compared with that due to changing steam rate.

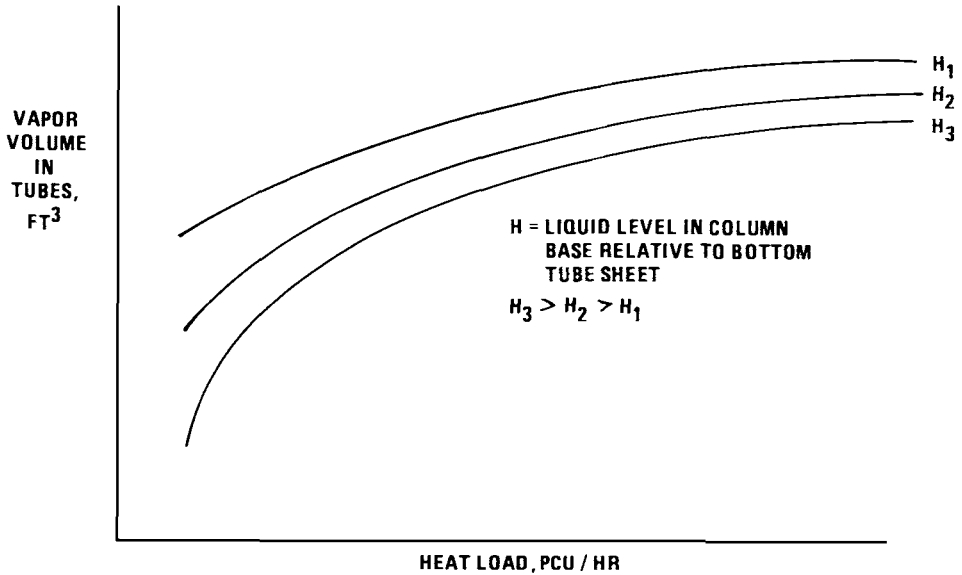


FIGURE 4.2

Relationship between vapor volume in tubes of thermosyphon reboiler, heat load, and base liquid level

E. Critical Versus Noncritical Steam Flow

When the steam valve pressure drop is high enough that flow is critical, reboiler shell pressure has no effect on steam flow rate. The reboiler has process dynamics different from those obtained with noncritical steam flow. For noncritical flow the steam flow or flow-ratio controller can have much higher gain (often by a factor of 5–10) than when steam flow is critical. If, as sometimes happens, the switching point from critical to noncritical flow occurs at a heat load close to normal operating heat load, controller tuning can present perplexing problems. If the controller is properly tuned for a higher load, the control loop may be unstable at a slightly lower load. Conversely, if the controller is tuned at a lower load, control may be very sluggish at higher loads. Preferably, therefore, the reboiler should be designed to operate over its normal range in one flow regime or the other.

F. High Concentration of Nonvolatile Components

Here the reboiler is fed with a wide boiling range mixture. For the case where the column-base liquid contains only a small amount of low boilers and a large amount of material with a much higher boiling point, perhaps approaching or exceeding maximum steam temperature, thermosyphon action and heat transfer may fall off as liquid level in the column base falls much below the elevation of the top tube sheet. What often happens is that either base level or low boiler concentration gets too low, thermosyphon action ceases, low boilers are completely stripped out in the tubes, and boiling stops. The column contents then drop into the column base. The increase in level and low boilers then permits boilup to resume. The alternate cessation and resumption are violent enough to merit the term *burping*. In view of what is often a rather narrow range of acceptable operating conditions, it is questionable whether a thermosyphon reboiler should be used for such applications. A forced-circulation reboiler would be better. If, however, a thermosyphon is used, one may maintain constant head on it by the overflow and external tank design of Figure 4.17.

4.3 FLOODED THERMOSYPHON (STEAM-SIDE) REBOILERS

As shown by Figure 4.3, a flooded thermosyphon reboiler operates by throttling condensate rather than steam flow. The basic principle is that of varying the heat-transfer surface. The chief advantage is that for a given steam flow in pounds per hour, a condensate valve is much smaller than a vapor supply valve. At a high heat load such that shell-side liquid level is low, condensate subcooling is small and there is some cavitation in or flashing across the condensate valve. Fortunately new methods developed by the ISA (see Chapter 11) permit much more accurate prediction of flashing and cavitation, and many valve manufacturers now can provide anticavitation trim. Care must be taken to avoid excessive reboiler ΔT , which may cause either film boiling or choked flow. A steam-supply pressure regulator is often used to protect against this.

Control of the condensate valve is sometimes achieved by a steam-flow measurement, sometimes by a condensate-flow measurement (more accurate and cheaper), and sometimes by another process variable. Response is more sluggish than when steam flow is throttled. No trap is required. The theory and design equations are presented in Chapter 15. For troubleshooting, performance tests, startup, and, in some cases, overrides, a liquid-level transmitter should be installed on the reboiler shell side.

A special version of a flooded reboiler designed for low-boiling materials (requiring low-temperature steam) is shown in Figure 4.4. Steam to the reboiler is throttled in a conventional fashion—usually flow controlled or flow-ratio controlled—but there is neither a trap nor a condensate pot. Instead condensate is removed through a loop seal whose top is vented to atmosphere. The height

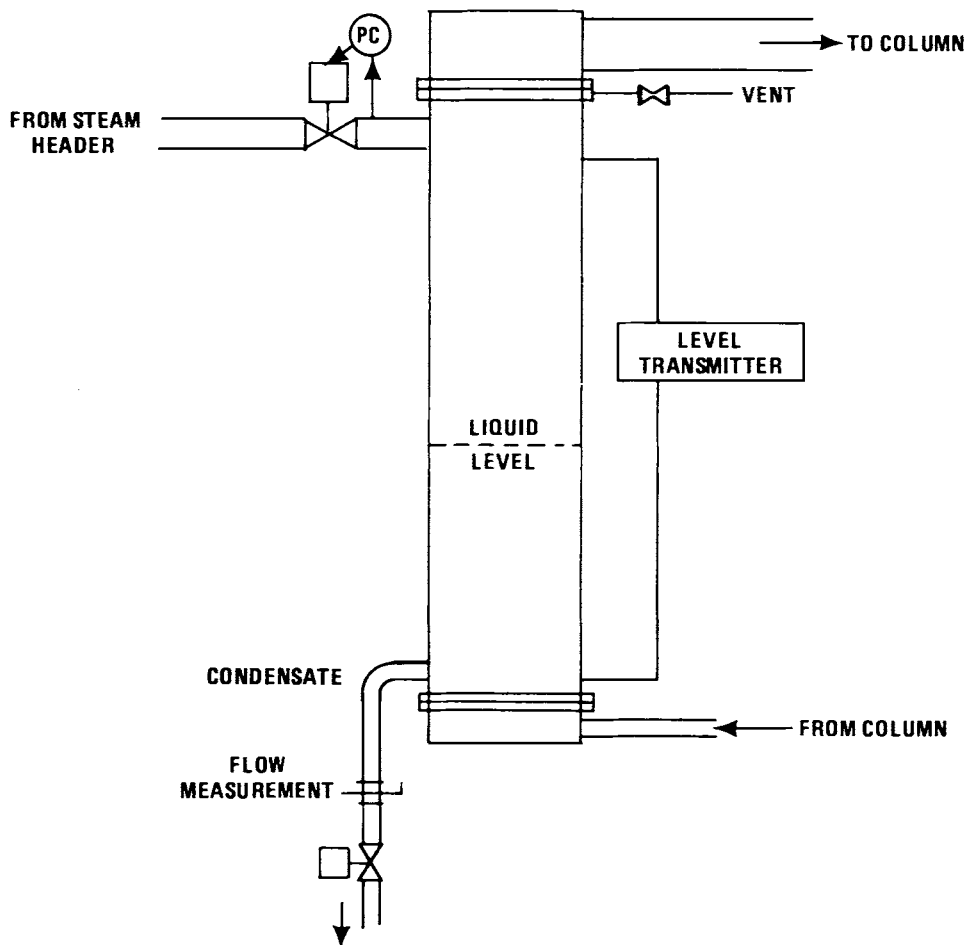


FIGURE 4.3
Flooded reboiler

of the loop is typically 5–10 feet. This arrangement has the equation:

$$H_L \rho \frac{g_L}{g_C} = H_S \rho \frac{g_L}{g_C} + P_S + \Delta P_{\text{line}} \quad (4.1)$$

where

H_L = loop-seal or standpipe height, feet

H_S = condensate level in the shell, feet

P_S = shell pressure, lbf/ft² absolute

ρ = condensate density, lbm/ft³

Now the absolute value of P_S does not change very much so boilup rate is mostly controlled by variation in exposed tube area.

4.4 FORCED-CIRCULATION REBOILERS

Forced-circulation reboilers are commonly horizontal as shown in Figure 4.5, but also occasionally are vertical. To minimize pressure drop, a restriction downstream of the reboiler is sized to prevent vaporization in the reboiler tubes. This restriction may be in the vapor line to the column or right at the vapor nozzle outlet.

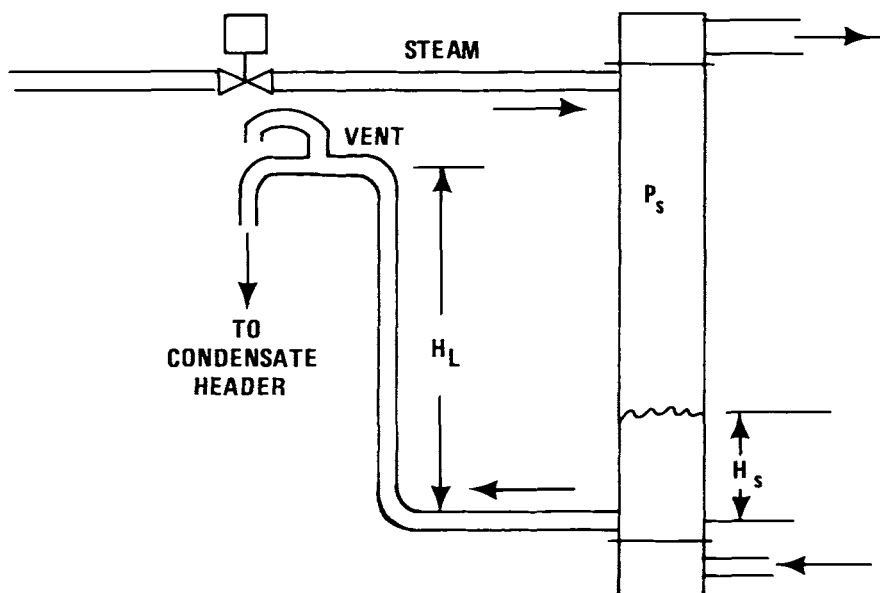


FIGURE 4.4
Flooded reboiler for low boiling point materials

Forced-circulation reboilers are widely used for vacuum columns because of their lower pressure drop, for applications where a thermosyphon would be expected to foul, and for applications where thermally sensitive materials are being distilled.

4.5 FLOODED-BUNDLE KETTLE REBOILERS

Kettle-type reboilers are sometimes used with vacuum columns. They eliminate the need for the circulating pump required with a forced-circulation reboiler, and avoid the temperature elevation encountered in the lower end of a thermosyphon reboiler. The usual arrangement is that of the flooded-bundle type

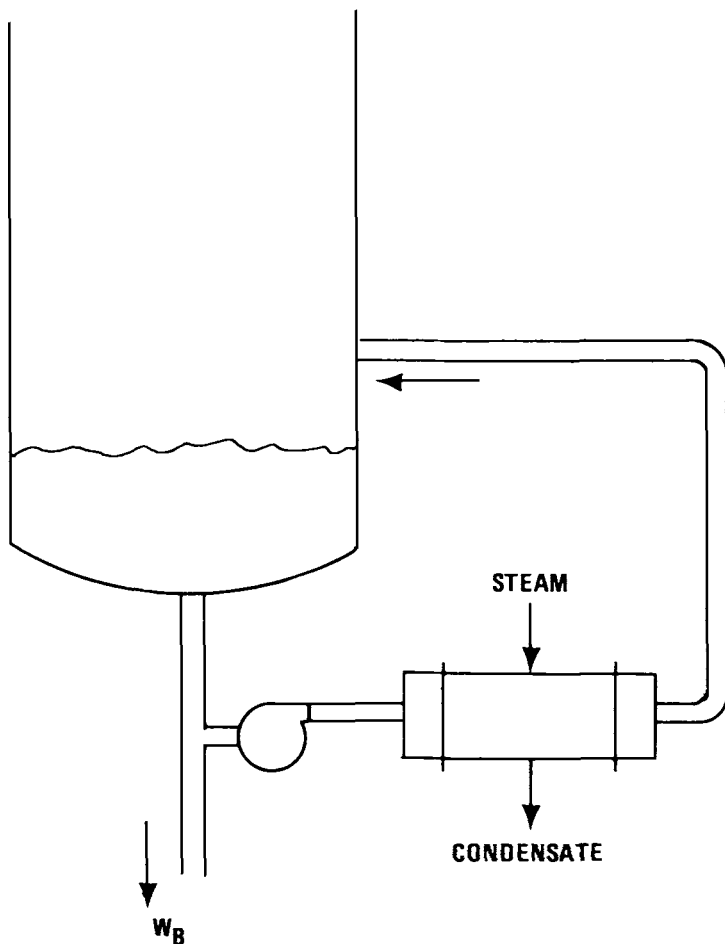


FIGURE 4.5
Column base with forced-circulation reboiler

shown on Figure 4.6. This is very similar to an internal reboiler with an isolating baffle or chamber, as will be discussed in the next section.

The column base runs empty and there is just enough liquid head in the line connecting the column base to the reboiler to overcome the pressure difference between the kettle reboiler and the column base. To protect the tube bundle in the event that boilup temporarily exceeds downflow, a head measurement must be made on the tube bundle chamber. This can be connected to an interlock, or, as shown on Figure 4.7, to suitable overrides.

Shown here are two overrides (pneumatic devices illustrated). One has a latch-up circuit with a gain 25 relay, which provides a high signal the first time the reboiler is filled with liquid; the steam valve is held closed until the level covers the tubes. Once latched up this circuit does not function again until it is unlatched by switching to "shutdown."

The other circuit is intended to shut the steam valve if the total head drops below a certain amount after normal operation has been in effect. This requires

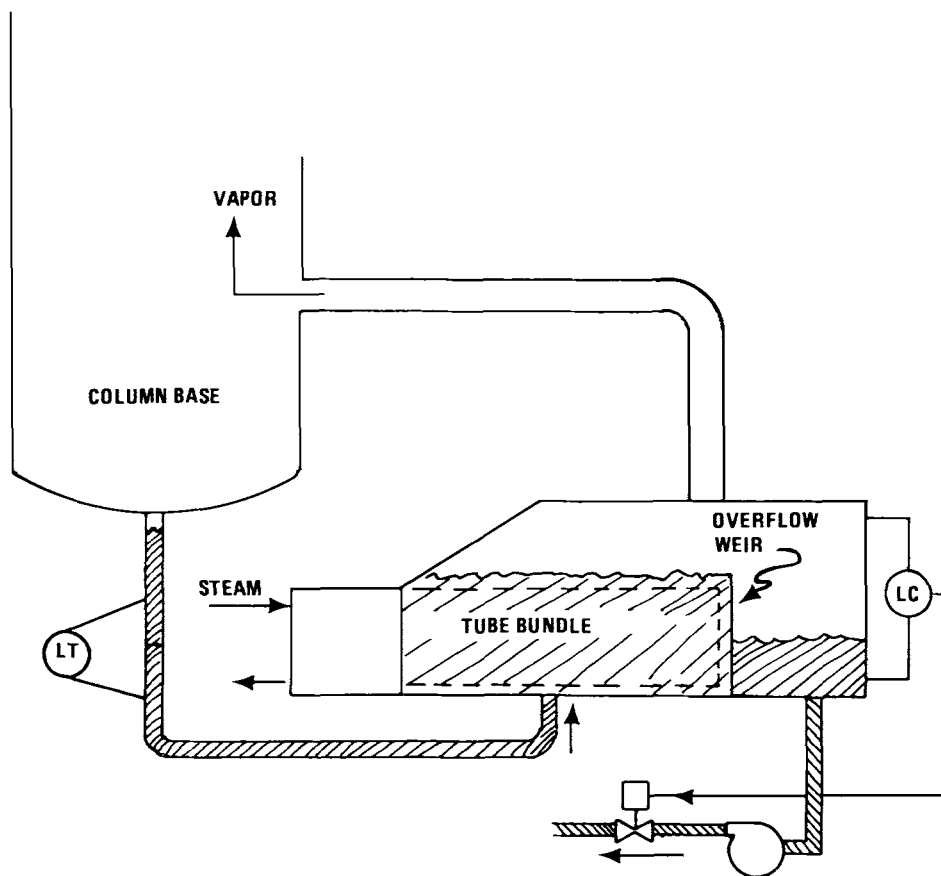


FIGURE 4.6
Kettle-type reboiler with internal weir

field calibration since the presence of froth between the tubes produces an average density well below that of clear liquid.

4.6 INTERNAL REBOILERS

Internal reboilers, usually in the form of horizontal, cylindrical bundles of U-tubes with the heating medium in the tubes, have been used to a modest extent. In comparison with external reboilers, pumping and pressure-drop considerations are eliminated, as are piping and reboiler shell. Offsetting these in part, however, is the fact that the column base must be longer (taller) to provide

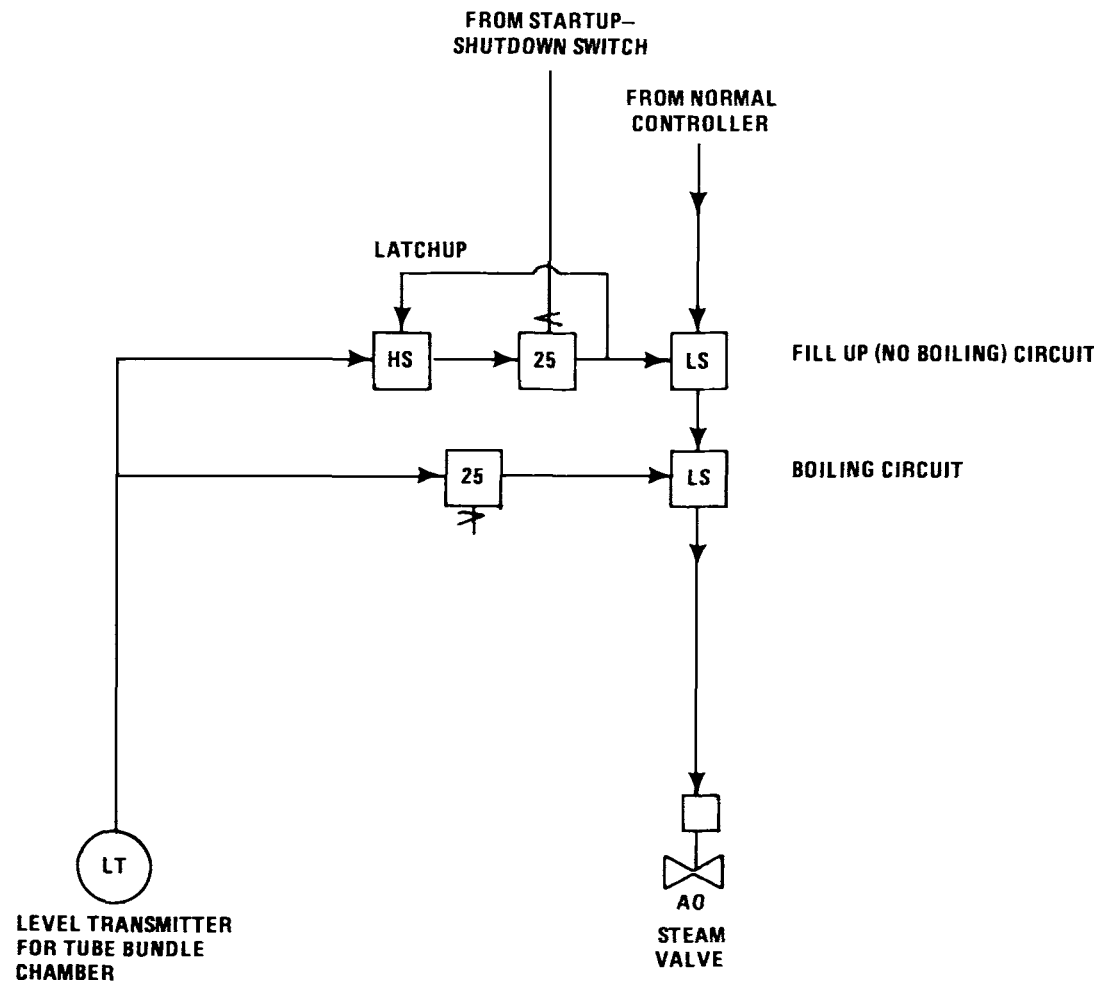


FIGURE 4.7
Protective circuits for tube bundle chamber in kettle-type reboiler

adequate space for controlling the froth level. There is also the added difficulty of controlling the level of a variable-density froth. In addition, the column must be shut down to clean the tubes.

Two basic arrangements for installing and operating internal reboilers are employed:

1. Tube bundle without isolating baffle or chamber.
2. Tube bundle with isolating chamber.

Tube Bundle Without Isolating Baffle or Chamber

This case is illustrated in Figure 4.8. Here the tube bundle* is submerged in a pool of liquid. Above the bundle we normally find froth, not clear liquid. Some bubbles, of course, exist in the spaces between the tubes. Since most level instruments really measure head, not level, there is the practical problem of measuring or predicting true level accurately enough to prevent the froth from rising high enough to flood the lower tray or trays. As indicated by the literature, this is no trivial problem.⁷

From the column or reboiler designer it is necessary to get an estimate of relative froth density:

$$\phi = \frac{\text{Froth density, lbm/ft}^3}{\text{Clear liquid density, lbm/ft}^3} \quad (4.2)$$

The Hepp⁷ correlation may be used to estimate ϕ :

$$\phi = 1 - 0.62 \, v \sqrt{\rho_v} \quad (4.3)$$

where

v = vapor velocity, ft/sec

ρ_v = vapor density, lbm/ft³

To get adequate holdup for level control, we must take ϕ into account. If, for example, $\phi = 0.5$, we need twice the volume that would be required for clear liquid. Thus this type of internal reboiler probably should not be used in vacuum towers since the required base froth volume would be too great.

For the particular case where $\phi = 0.5$, the nozzle spacing of Figure 4.8 is suggested. Note that L_1 and L_3 are the nozzles used for level control (see discussion in Chapter 11). L_3 and L_2 are used to determine that the bundle is covered by liquid. To allow for variations in froth density, the top 24 inches between L_2 and L_3 are clearance; the level transmitter is calibrated such that "100 percent" level is 76 inches above L_2 .

Although liquid level can come a little way below the top of the tube bundle without causing either fouling or a reduction in heat transfer, it is more conservative to keep the bundle submerged.

* Sometimes the bundle consists of U-tubes; sometimes it is a two-pass bundle extending through the column with the intermediate head on the far side. The latter design is better than the former for flooded-tube operation.

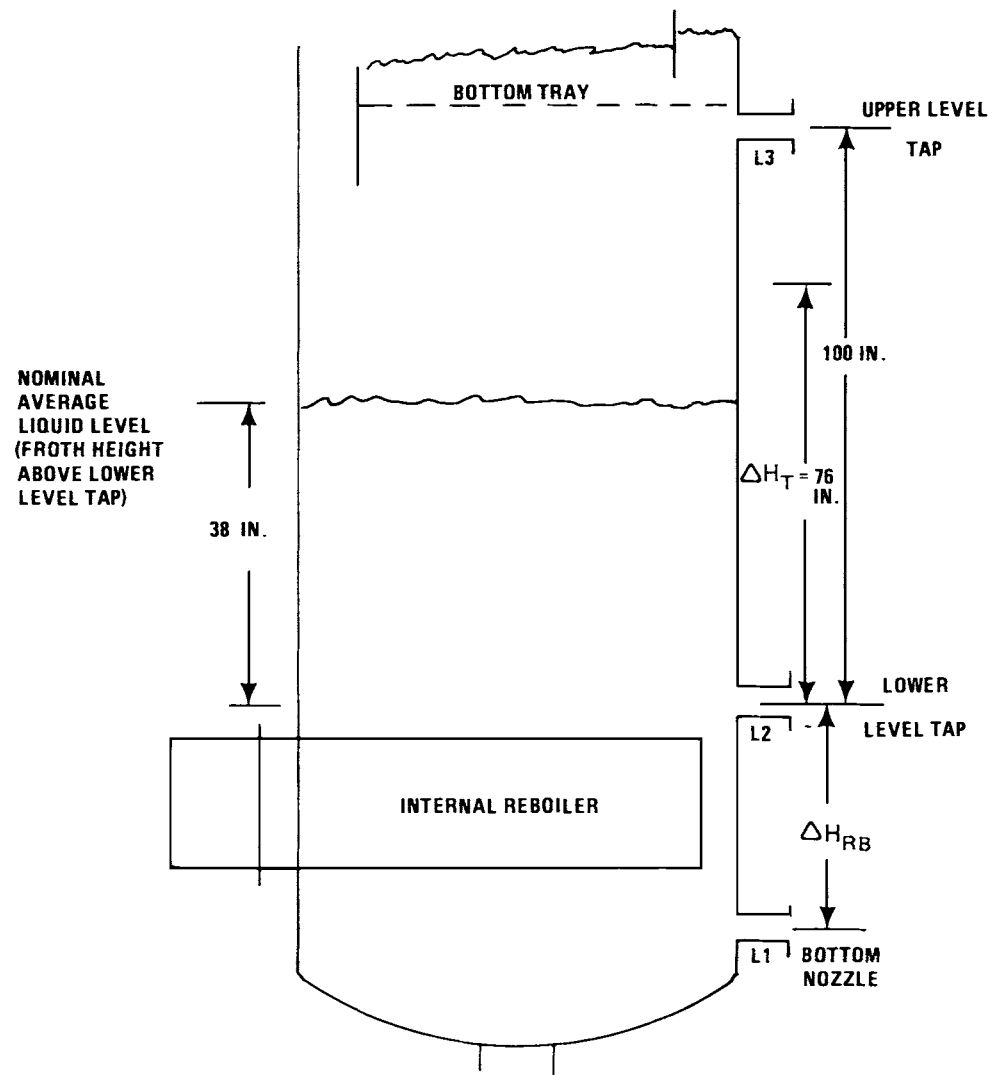


FIGURE 4.8
Column base with internal reboiler—no baffles or weirs

Tube Bundle with Isolating Chamber

As shown by Figure 4.9, internal reboilers are sometimes isolated from the column base by an isolating baffle, chamber, or "bathtub." The downcomer from the last tray dumps into the chamber and the excess overflows into the column base. The outlet weir from the chamber is generally set high enough to flood the tube bundle. There are now two problems: (1) controlling level in the column, and (2) protecting the tube bundle if boil-off exceeds downflow.

In theory it is possible to run base level at elevations above the bottom of the isolating chamber (in the space between L_2 and L_4 in Figure 4.9). Some columns have been built this way. In practice it is undesirable because cross-sectional area, and therefore holdup at this elevation, are limited. Thus it is recommended that the level be measured and controlled below the isolating chamber.

In the event that boilup exceeds downflow, the level in the isolating chamber drops and exposes tubes. With many organics, this causes a type of fouling called "varnish." To prevent this from happening, one should provide a second level transmitter, connected to nozzles L_3 and L_4 . The output of this transmitter should be connected to an override in the heating medium control valve circuit; a low level should pinch this valve.

4.7 STEAM SUPPLY AND CONDENSATE REMOVAL

Steam supplies for distillation columns are commonly arranged as shown in Figure 4.10. From a high-pressure header, steam is let down to a lower pressure through a "reducing station." This may feature a transmitter, controller, and valve, but is often a self-contained, self-actuated pressure regulator. In some cases there is also a desuperheater. This last item is sometimes installed because of a common misunderstanding about the effect of superheat on reboilers; it simply reduces slightly the amount of condensate⁸ and usually does not, as sometimes believed, reduce reboiler heat-transfer capacity.

The reduced-pressure header may serve one load or many. For a design with only two or three loads, there is often a severe interaction between the loads and the pressure reducer. Suppose, for example, that steam flow to one reboiler is increased. This drops pressure in the reduced-pressure header. The pressure controller opens the upstream valve to restore pressure, but in the meantime the lower pressure may have caused another reboiler supply valve to open. If all of the controllers are tuned tightly, there may ensue "fighting" among the controls and substantial swings in header pressure and steam flows to the various reboilers.

For a header with many loads, header pressure is really constant at only one point; at the far end of a distillation train, header pressure may be much lower. Particularly for a small number of loads, it may be more logical to eliminate the pressure reduction. A system such as shown in Figure 4.11 may

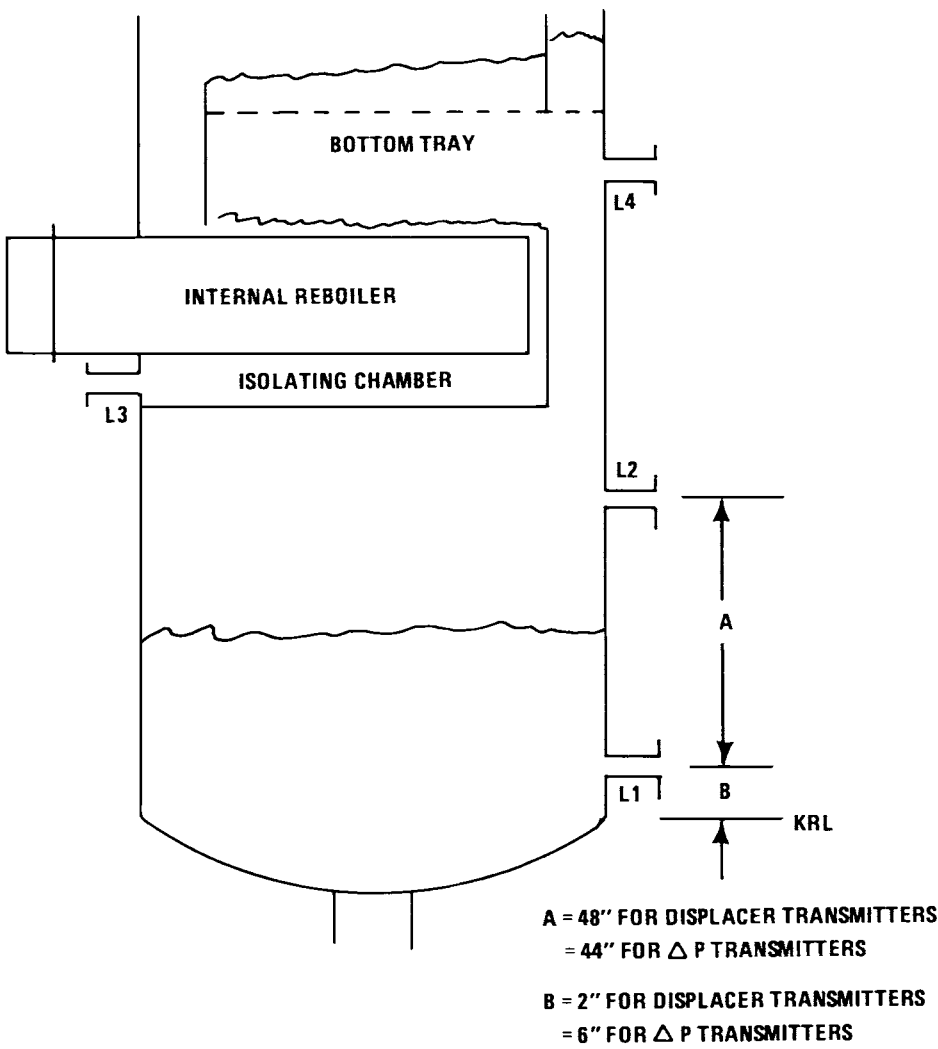


FIGURE 4.9
Column base with isolated internal reboiler

be used instead. Here each reboiler is supplied from a valve connected directly to the high-pressure header. This saves one valve and controller and the higher pressure drop means smaller valves. There is more likelihood of critical pressure drop, and less probability that flow regimes will switch during normal operation (see Section 4.2). This figure also illustrates temperature and pressure compensation of the steam flow meter. Header pressure and temperature variations severely impair the accuracy of many uncompensated plant steam flow meters. A suitable scheme for temperature and pressure compensation is given in Chapter 11. Errors in steam-flow metering without such compensation are also discussed.

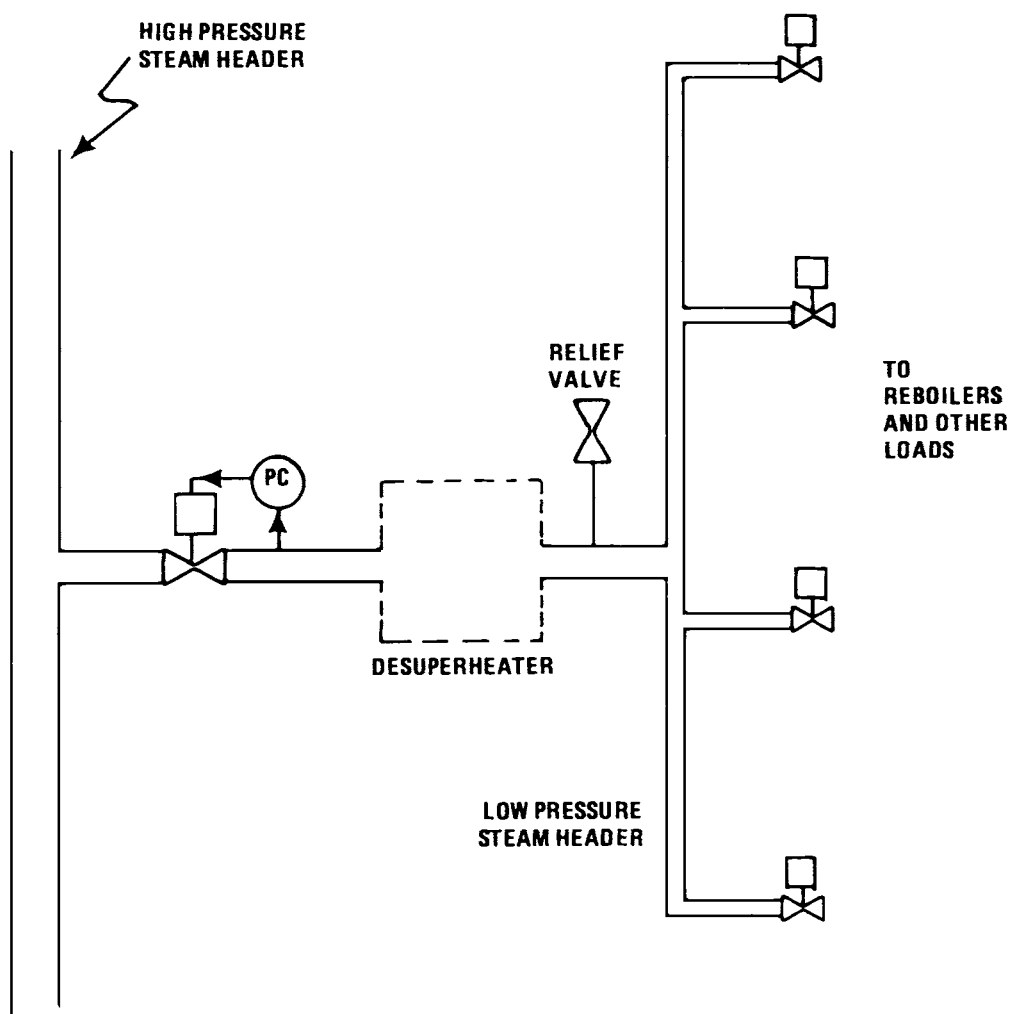


FIGURE 4.10
Steam header configuration

When header pressure variations are small, temperature compensation is sometimes omitted since temperature changes more slowly than pressure.

Steam condensate removal is most commonly achieved by traps—devices that, in effect, are very simple level controllers. Many plants have experienced the need for high maintenance with these. Sometimes enough steam leaks through to impair the accuracy of steam-consumption estimates based on steam flow metering. In addition, as pointed out by Mathur,⁹ a rapid closure of the steam valve may cause a vacuum in the shell and pull back condensate through the trap with possible hammering and vibration in the shell. As a result, some reboilers, particularly large ones, are now equipped with condensate seal pots. These pots then have conventional level controllers.

Two practical problems are associated with the selection of condensate removal valves: (1) in sizing calculations allowance usually must be made for some flashing, and (2) the pressure drop increases with flow rate since reboiler shell pressure increases with load. This means that the installed valve flow

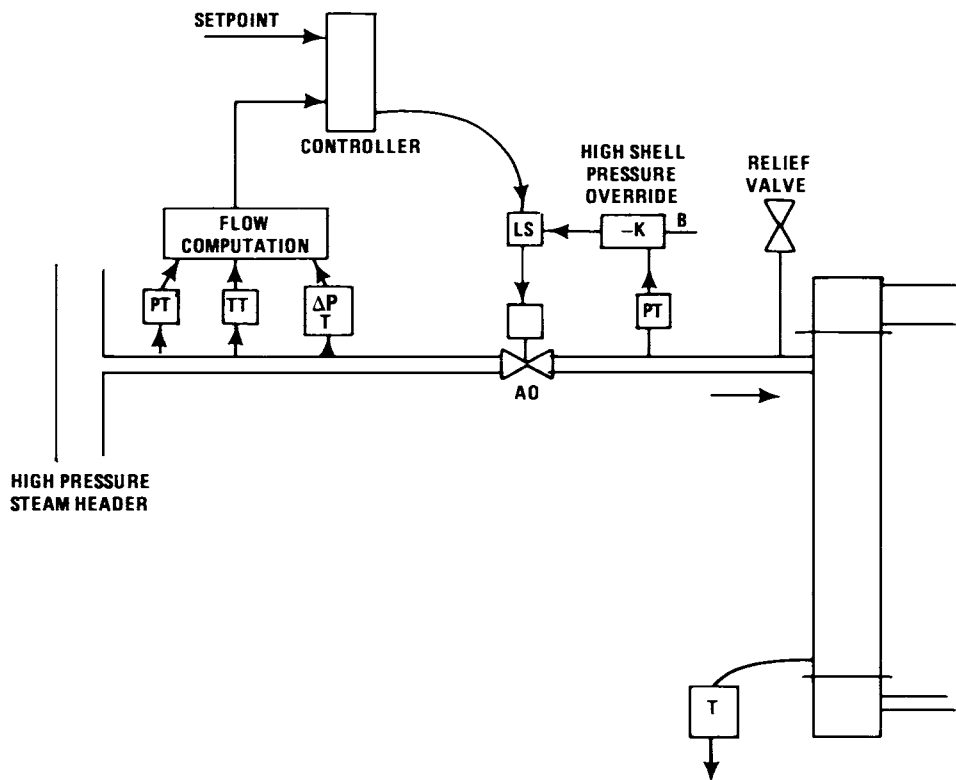


FIGURE 4.11
Improved steam supply and flow control system

characteristic should be between linear and square root. Anticavitation trim may be required.

4.8 REQUIRED HOLDUP FOR LEVEL CONTROL

Base level is most commonly controlled by manipulating one of three variables: (1) bottom-product withdrawal, (2) steam or other heating medium flow, or (3) feed flow. Avoid scheme 2, if at all possible, because of reboiler and column-base "swell," mentioned earlier. It is sometimes required, however, if the bottom-product flow is very small.

In the discussion that follows, the basic criterion for choosing column-base holdup will be good, or at least adequate, control of the column. As mentioned earlier, good composition control is favored by small holdups. But if the column base serves as a feed vessel for another step, its required volume may be influenced by downstream requirements. This is discussed in Chapter 5. If possible, we will avoid the use of separate, intermediate buffer tanks to feed another process step.

For maximum flow smoothing, we recommend a PI level controller cascaded to flow control. The PI level controller should have, as indicated by Figure 4.12, auto overrides to keep the liquid level between the level transmitter taps. Quantitative design is discussed in Chapter 16.

Base-Level Control via Bottom-Product Withdrawal

In addition to the auto overrides of Figure 4.12, we commonly have three more overrides. One pinches the bottom product flow if a downstream level gets too high. A second pinches the feed flow if column base level becomes too high, while a third pinches bottom product flow if column base composition has too high a concentration of low boilers.

The characteristic time constant for this system is:

$$\tau_H = \frac{A_B \Delta H_T}{K_{cb} Q_{mFB}} \quad (4.4)$$

where

A_B = column-base cross-sectional area, ft²

ΔH_T = level transmitter span, feet

K_{cb} = level controller gain

Q_{mFB} = bottom-product flow-transmitter span, ft³/min

For this system it is usually satisfactory to make $\tau_H \geq 10$ minutes.

If flow smoothing is noncritical (as, for example, if bottom-product flow goes to a large storage tank), it will be satisfactory to use a proportional-only level controller as shown in Figure 4.13.

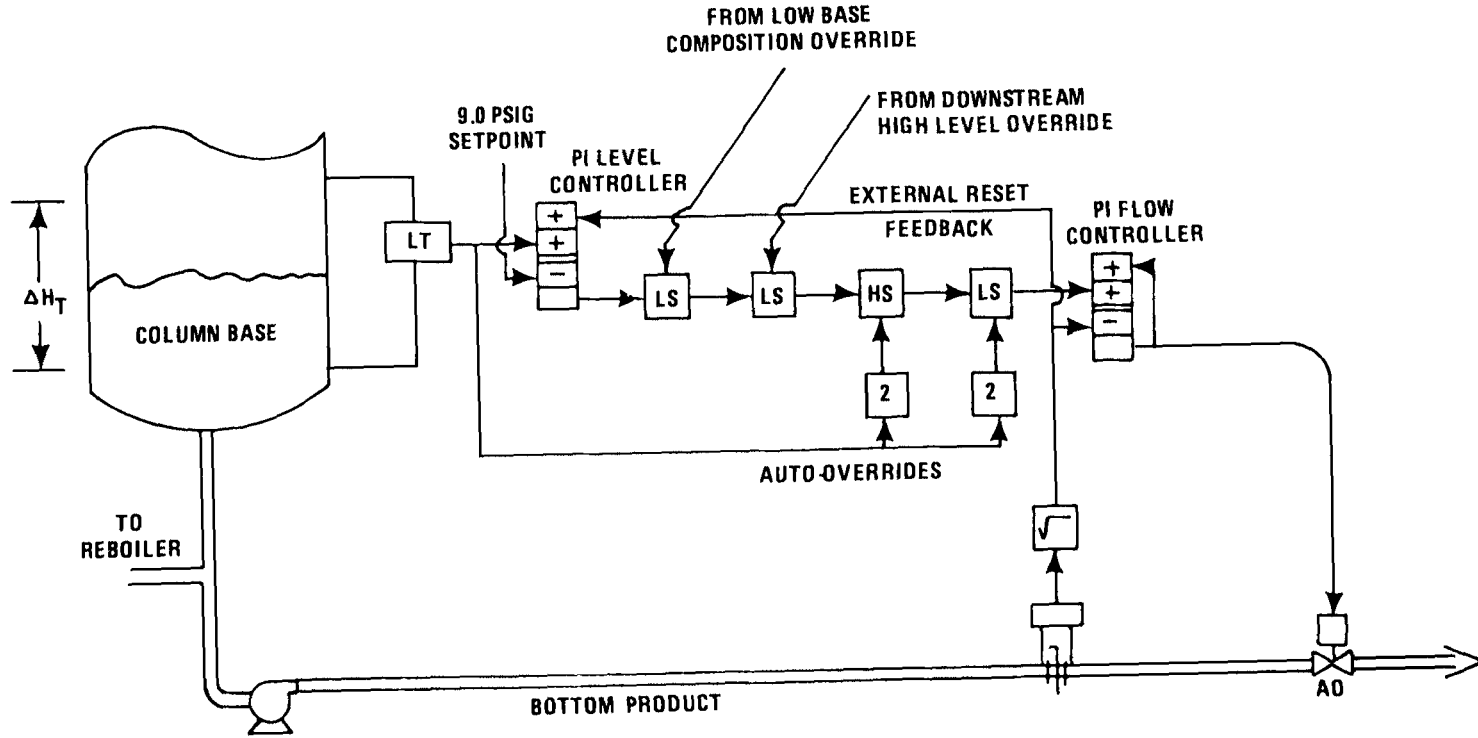


FIGURE 4.12
Level control of column base via bottom product throttling, cascade arrangement

The dynamic response of the proportional-only, averaging level control system may be defined by a first-order time constant:

$$\tau_H = \frac{A_B \Delta H_T}{K_c \times 12 \times \frac{dQ_o}{d\theta_c}} \quad (4.4a)$$

where

τ_H is in minutes

A_B = cross-sectional area, ft^2 , of the column base or vaporizer separator

ΔH_T = level transmitter span, in feet, of process fluid

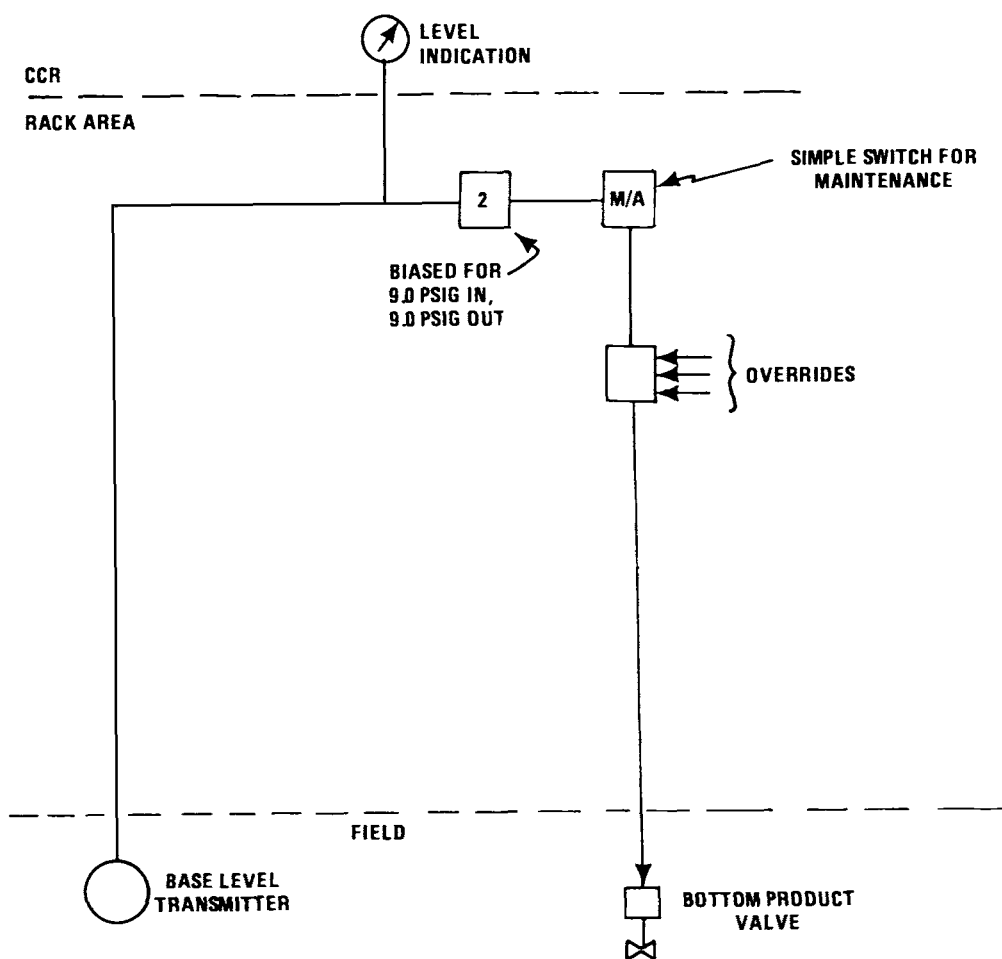


FIGURE 4.13

Proportional-only level control system for column base

K_c = controller gain

Q_o = manipulated flow, ft³/min

θ_c = signal to valve positioner, psig

For a linear installed flow characteristic,

$$\frac{dQ_o}{d\theta_c} = \text{valve gain} = (Q_o)_{\max}/12$$

$$(Q_o)_{\max} \cong 4 Q_{FS}$$

$$Q_{FS} = \text{flow sheet or average flow, ft}^3/\text{min}$$

There is some uncertainty in valve gain because of differences in the valve-sizing philosophy followed by various instrument engineers. As a result of some adverse experiences due to undersizing holdups by using a factor of 2 to multiply Q_{FS} to get $(Q_o)_{\max}$, we now lean to using a factor of 4 for design calculations early in a project. See, however, the discussion in Chapter 11.

Earlier it was suggested (Section 4.1) that a generally useful level transmitter span would be 38 inches of process fluid. Then if the valve positioner input span is 12 psi:

$$A_B = \frac{2 \times 4 \times A_{FS}}{38/12} \times \tau_H \quad (4.5)$$

$$= 2.53 Q_{FS} \tau_H \quad (4.6)$$

This is the required column-base cross-sectional area.

Base-Level Control via Feed-Flow Manipulation

Here τ_H should be 20 minutes or more and base-level control should be cascaded to feed-flow control. A mathematical analysis is given in Chapter 16.

Base-Level Control via Steam-Flow Manipulation

In this case τ_H should be 20 minutes or more. In addition, the level controller should be cascaded to a steam-flow controller with a linear flow transmitter (or orifice ΔP transmitter, and square-root extractor). Further, for a thermosyphon reboiler, one should make the volume $A_B \times \Delta H_T$ at least ten times the volume inside the reboiler tubes.

For this design:

$$\tau_H = \frac{A_B \Delta H_T}{K_{cb} \left[\frac{(w_{\sigma})_m \lambda_{\sigma}}{\rho_p \lambda_p} \right]} \quad (4.7)$$

where

$(w_{\sigma})_m$ = input span of steam flow transmitter, pounds per minute; that is, maximum steam flow

λ_{st} = steam latent heat of condensation, pcu/lb

λ_p = process fluid latent heat of condensation, pcu/lb

ρ_p = process fluid density, lbm/ft³

Note that the term in brackets

$$\frac{(w_{st})_m \lambda_{st}}{\rho_p \lambda_p}$$

is the process vapor flow corresponding to $(w_{st})_m$. A typical control system arrangement for level control via steam-flow control is shown in Figure 4.14.

Much difficulty has been encountered when trying to control base level by adjusting steam flow to a thermosyphon reboiler. It therefore is recommended that all such applications be subjected to a dynamic analysis such as presented in Chapter 16.

Minimum Holdup Design

For those distillations involving thermally degradable materials, it is necessary to keep the amount of hot-liquid storage to a minimum. The preferred design of Figure 4.15 involves maintaining liquid level in a small-diameter pot or "peanut." Level nozzle spacing should be chosen as before for the same reasons. If a thermosyphon reboiler is used, care should be taken to ensure that the lower level nozzle is not below the midpoint of the tube length. Note that for overall process control, it may be necessary to send the bottom product to a cooler and surge tank to get adequate flow smoothing for the next process step. (See Figure 4.17.) If this arrangement is used, the bottom product valve may be used with an override to maintain a low level in the column base. When this valve is pinched, the bypass valve opens to send material to the column base.

If a cooler is used, it should be noted that the surge tank may act as a condenser via the equalizing line. This may or may not be undesirable.

4.9 MISCELLANEOUS COLUMN-BASE DESIGNS

A large number of column-base designs have appeared in the literature,¹⁰ and it is our understanding that there are various proprietary designs that have never been publicized. Many of these include baffle arrangements intended to make the column base and reboiler function as an efficient stage of separation. An occasionally encountered design for this purpose is illustrated in Figure 4.16. As can be seen, the thermosyphon reboiler suction is taken from behind the outlet weir for the last downcomer, which extends almost to the base. This means that reboiler feed is richer in low boilers than is the column-base contents.

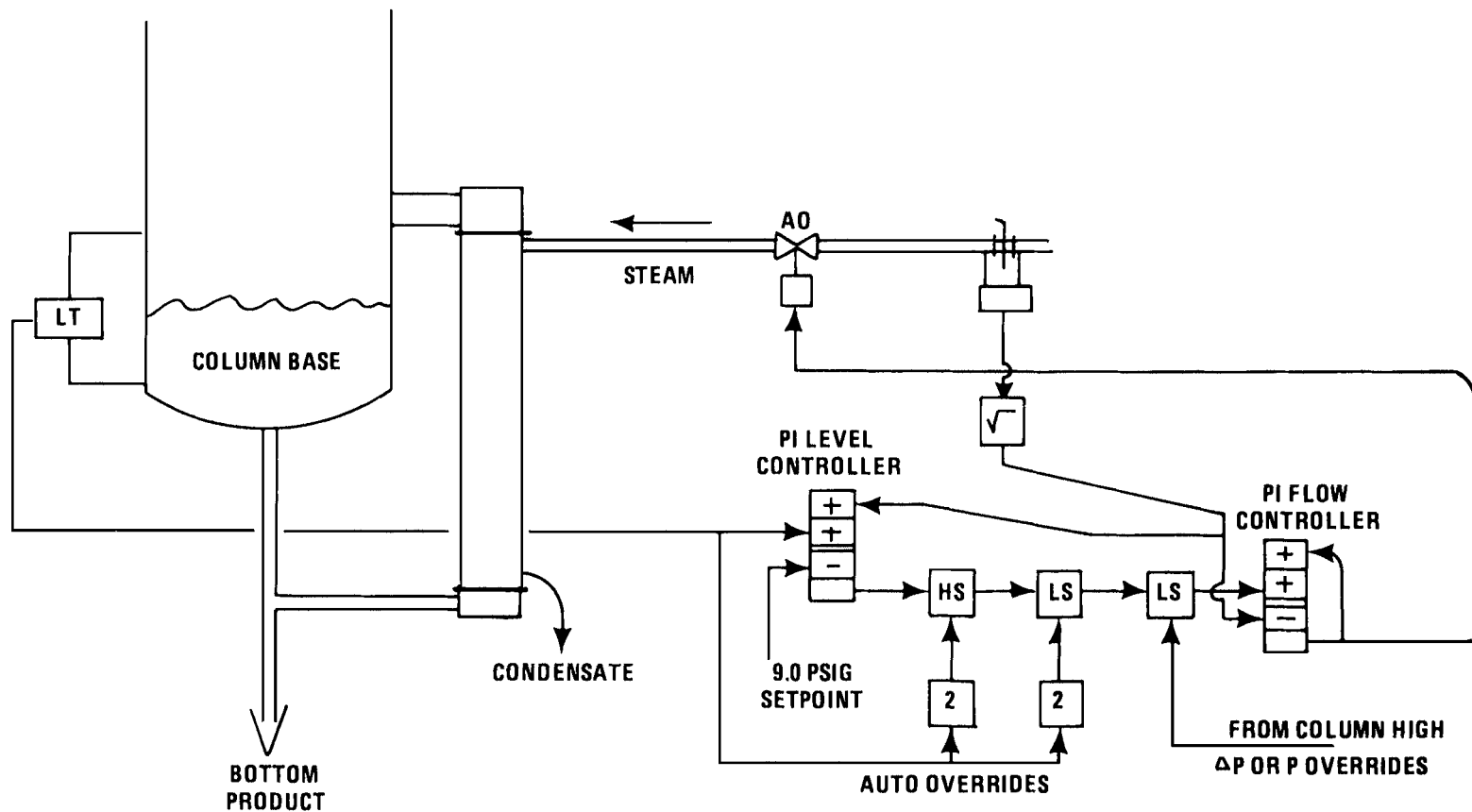


FIGURE 4.14
Column base level control by steam flow manipulation, cascade arrangement

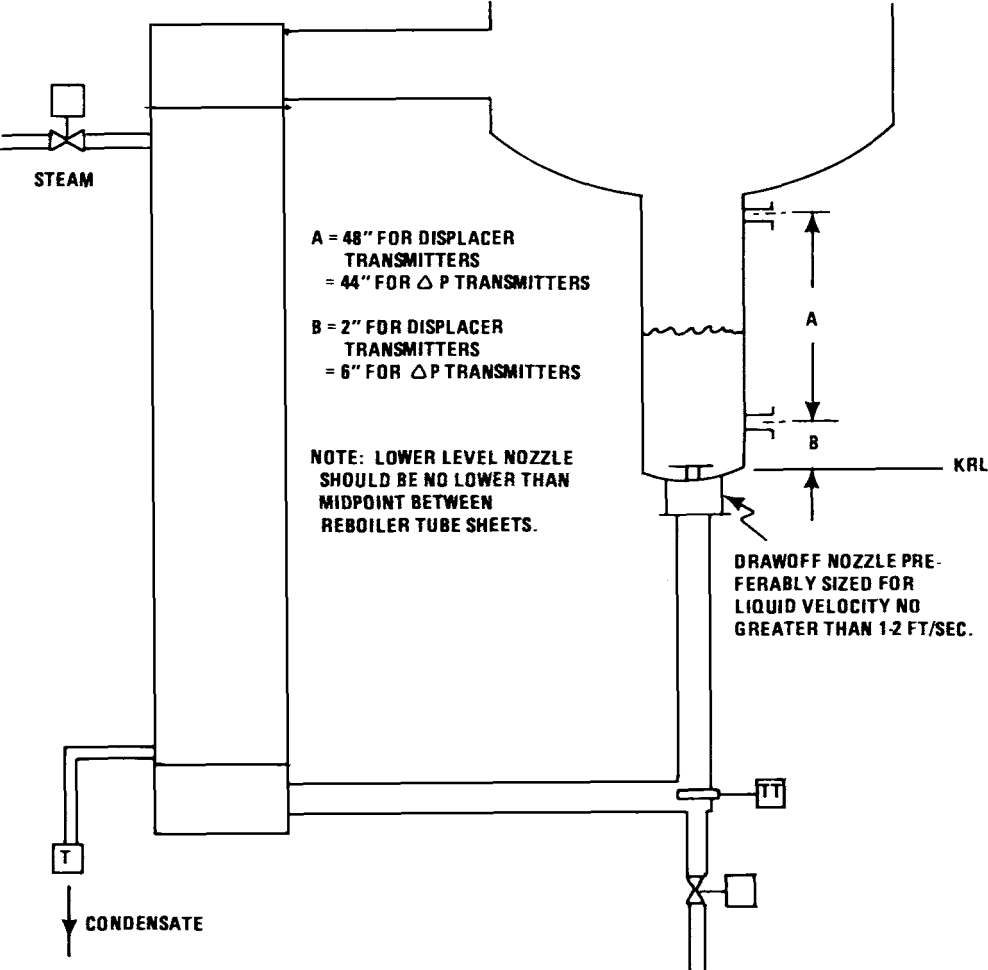


FIGURE 4.15
Column base design and arrangement for minimum holdup

When the column base is too small for smooth level control, an external surge tank must be used. One way of connecting the two is to overflow from the column base into the bottom product receiver as shown in Figure 4.17. A recycle line back to the column base (or to a feed tank) facilitates startup and can be used to protect the column base—and therefore also the reboiler—from excessively low level. The bottom product receiver, as a minimum, should have enough holdup to contain the entire column contents.

4.10 MISCELLANEOUS REBOILER DESIGNS

Two other types of reboilers are sometimes encountered: (1) reboilers with hot oil as a heating medium, and (2) direct-fired reboilers. Both are more common in the petroleum industry than in chemical plants.

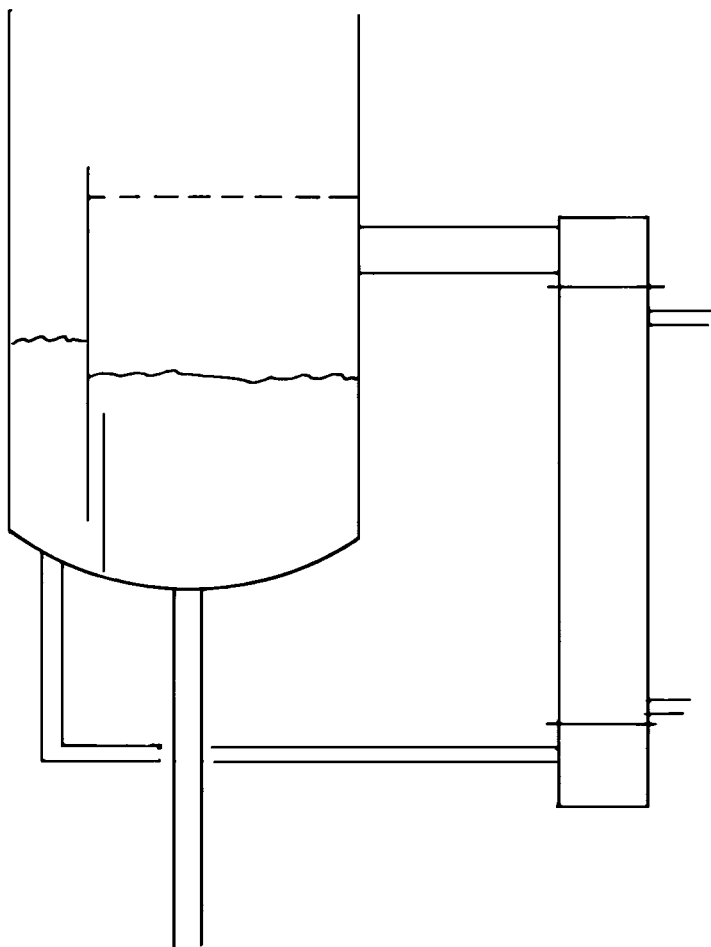


FIGURE 4.16
Reboiler piping arrangement for preferential boiling of reflux from lowest downcomer

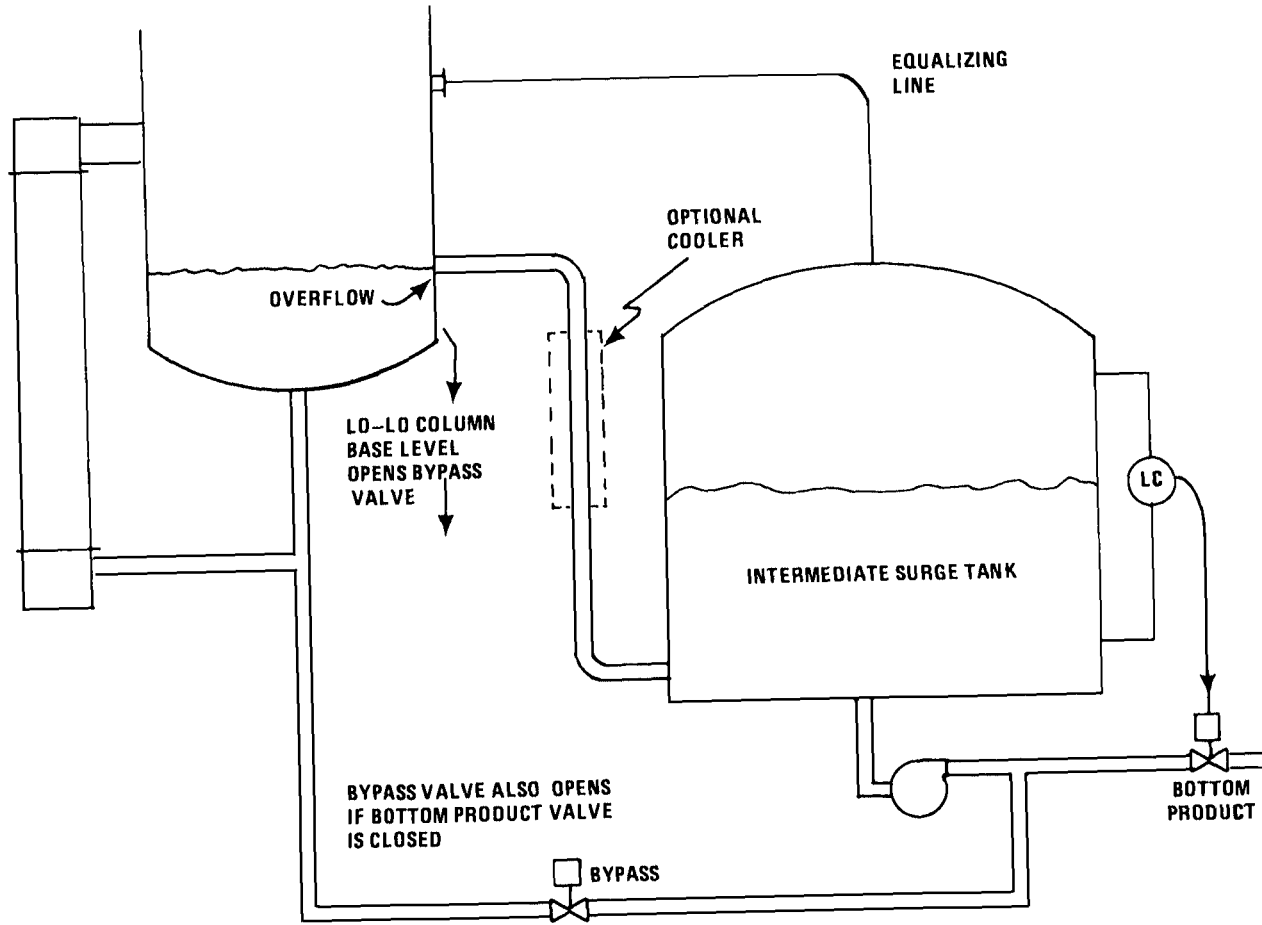


FIGURE 4.17
Arrangement for column base overflow into intermediate surge vessel

Hot-oil reboilers may be like steam-heated reboilers except that there is no phase change on the oil side. Best practice is to estimate boilup w_{BU} from the equation:

$$w_{BU} \lambda_{BU} = w_{oil} c_p (T_{in} - T_{out}) \quad (4.8)$$

This requires measuring oil-flow rate and inlet and outlet temperatures.

For higher temperatures than available with steam, use is made of other condensing media such as Dowtherm or p-cymene. For even higher temperatures, direct-fired reboilers may be used.

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5

Feed System Arrangements

5.1 GENERAL COMMENTS

the feed system for a column should function as a filter for incoming disturbances in feed flow rate, feed composition, and sometimes feed enthalpy. For minimum energy consumption operation, it should also send the feed to the proper feed tray. And for startup/shutdown of the column being fed, it may serve to receive recycled column product streams. In the following discussion, we will assume that (1) process material-balance control is in the direction of flow, (2) feed-tank level control is of the averaging type, and (3) there is good mixing in the tank.

5.2 FEED FLOW CONTROL

Figure 5.1 shows three commonly encountered feed flow schemes. When the upstream pressure is higher than the column pressure, then only a letdown valve is required, as shown by Figure 5.1A. If column pressure is greater than upstream pressure, then a pump is required, as shown by Figure 5.1B. If, however, upstream or downstream pressures can vary significantly, then a cascade level control/liquid flow control system such as that of Figure 5.1C is required. The flow signal should be linear—one should use a linear flowmeter or a square-root extractor with an orifice and ΔP transmitter. This (Figure 5.1C) is really the preferred overall design; it provides the most protection and offers the operator the maximum flexibility.

For small feed rates, it may be considerably cheaper to use a positive-displacement pump (piston type). Control may be by stroke adjustment on a constant-speed pump as shown by Figure 5.2A or by adjustment of a variable-speed drive of a pump with a fixed stroke as shown by Figure 5.2B. If the minimum stroke rate is at least three times the reciprocal of the feed-tray holdup time, no pulsation damper is required.

Some care should be taken in locating the feed control valve if flashing can occur. Hydraulic problems have been experienced in columns where the feed

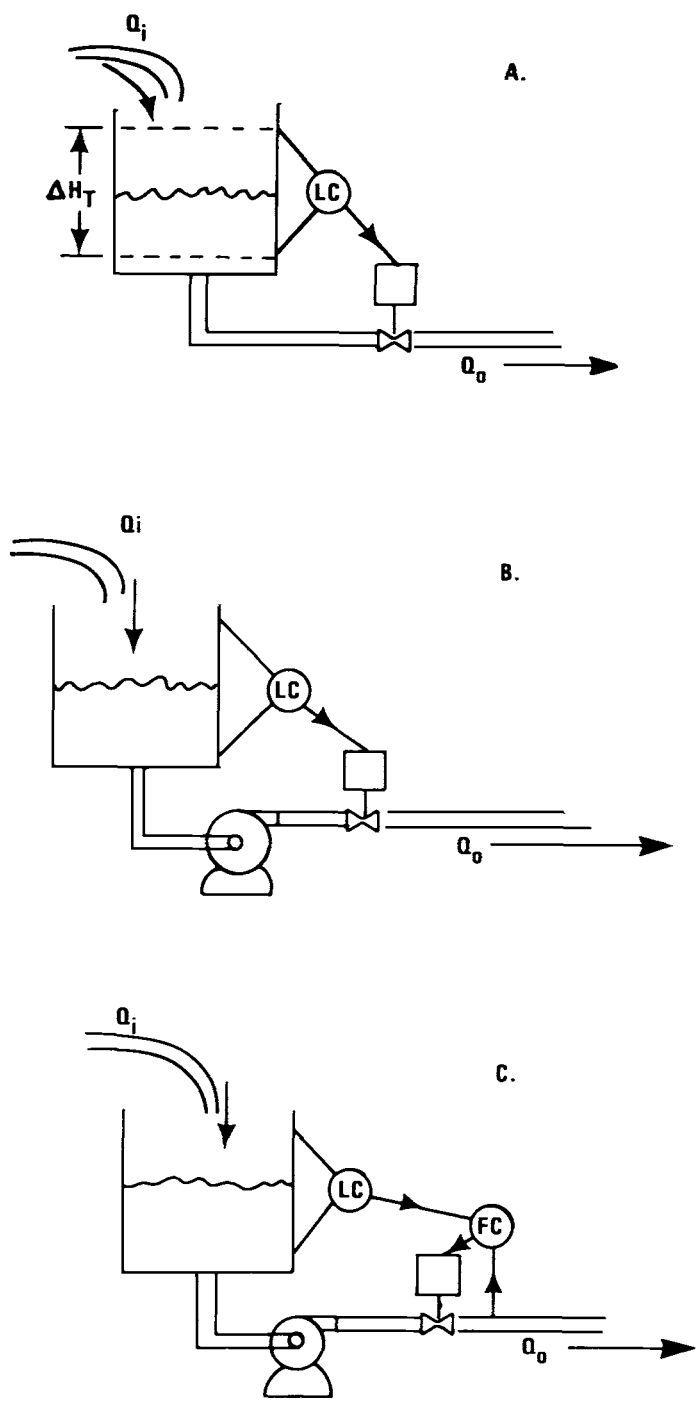


FIGURE S.1
Feed system for distillation column

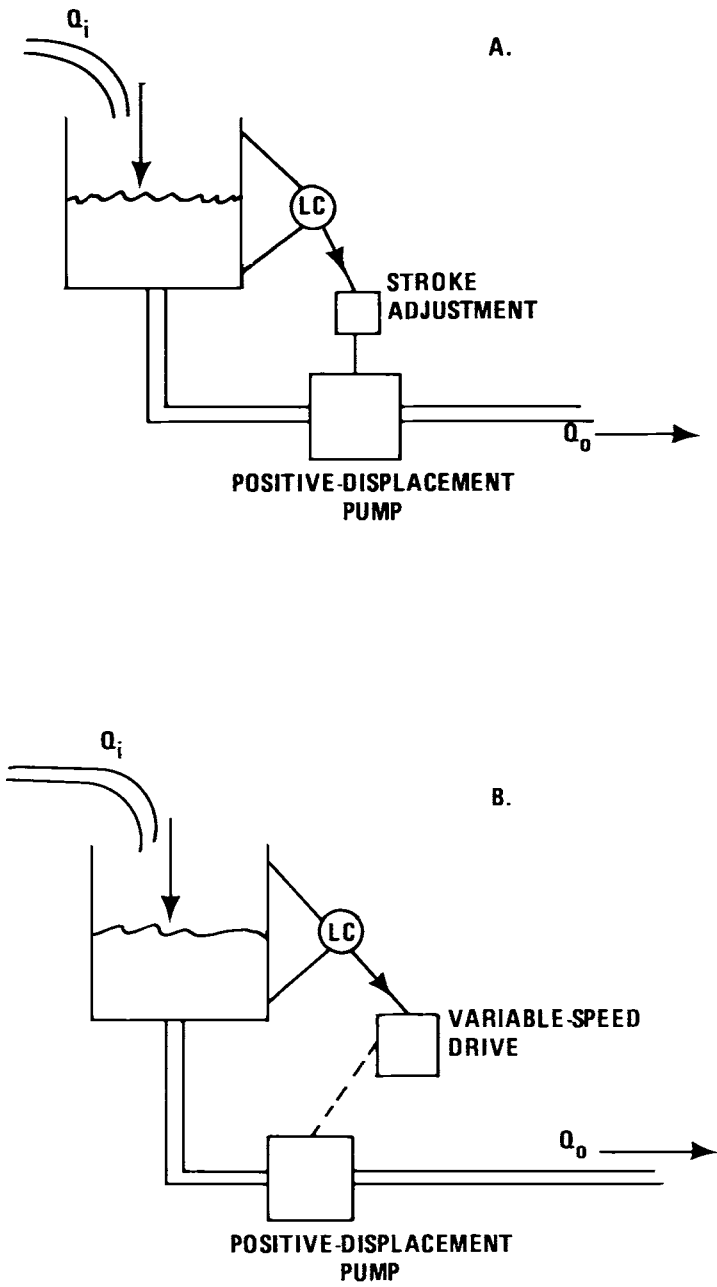


FIGURE 5.2
Column feed systems with positive-displacement pumps

valve is located at a lower elevation than the feed tray. Slugs of liquid can be dumped intermittently onto the feed tray as vapor slugs push up through the feed line. Therefore, where feed flashing can occur, the feed valve should be located as close as possible to the column inlet feed nozzle.

5.3 FEED TEMPERATURE CONTROL

Fluctuations or variations in the column heat balance can be a major factor in interfering with good composition control. One source of heat disturbances is the column feed, particularly if a heat economizer such as that shown in Figure 5.3 is used. The objective—to minimize heat consumption—is a worthy one, but this system will not, without a bypass, provide constant feed temperature or enthalpy. With a bypass, one can achieve temperature control.

Accordingly this arrangement is occasionally modified by the addition of a heater. Shown in Figure 5.4 is such a system with a temperature control scheme

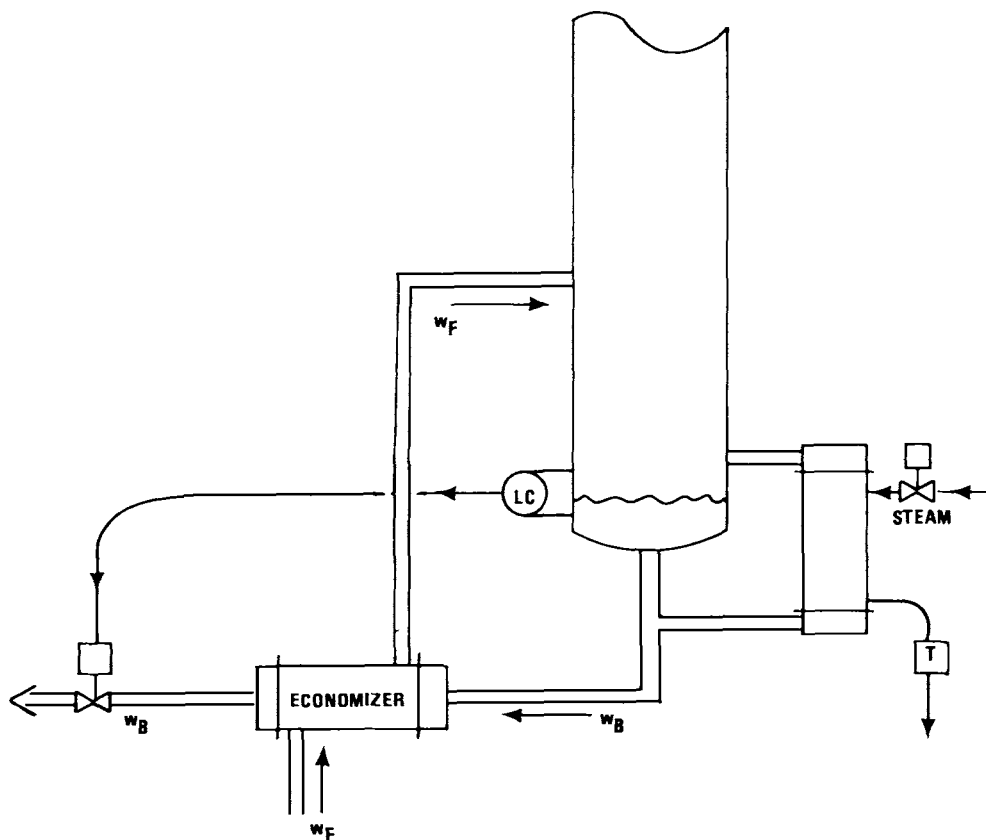


FIGURE 5.3
Column feed preheat via exchange with bottom product

that holds feed temperature constant by throttling steam flow (hot oil is sometimes used). This fixes the feed enthalpy if the feed is not partially vaporized (see the next section for a discussion of feed enthalpy control).

Column feed is sometimes preheated in other heat-exchanger systems, some of which are quite complex. Overhead vapor from the column, sidestreams (liquid and, particularly, vapor), and other process streams are sometimes used to minimize energy consumption.

5.4 FEED ENTHALPY CONTROL

If the feed is to enter at its bubble point, or if it is partially vaporized, then we need an enthalpy control scheme such as that shown in Figure 5.5.^{1,2} Since

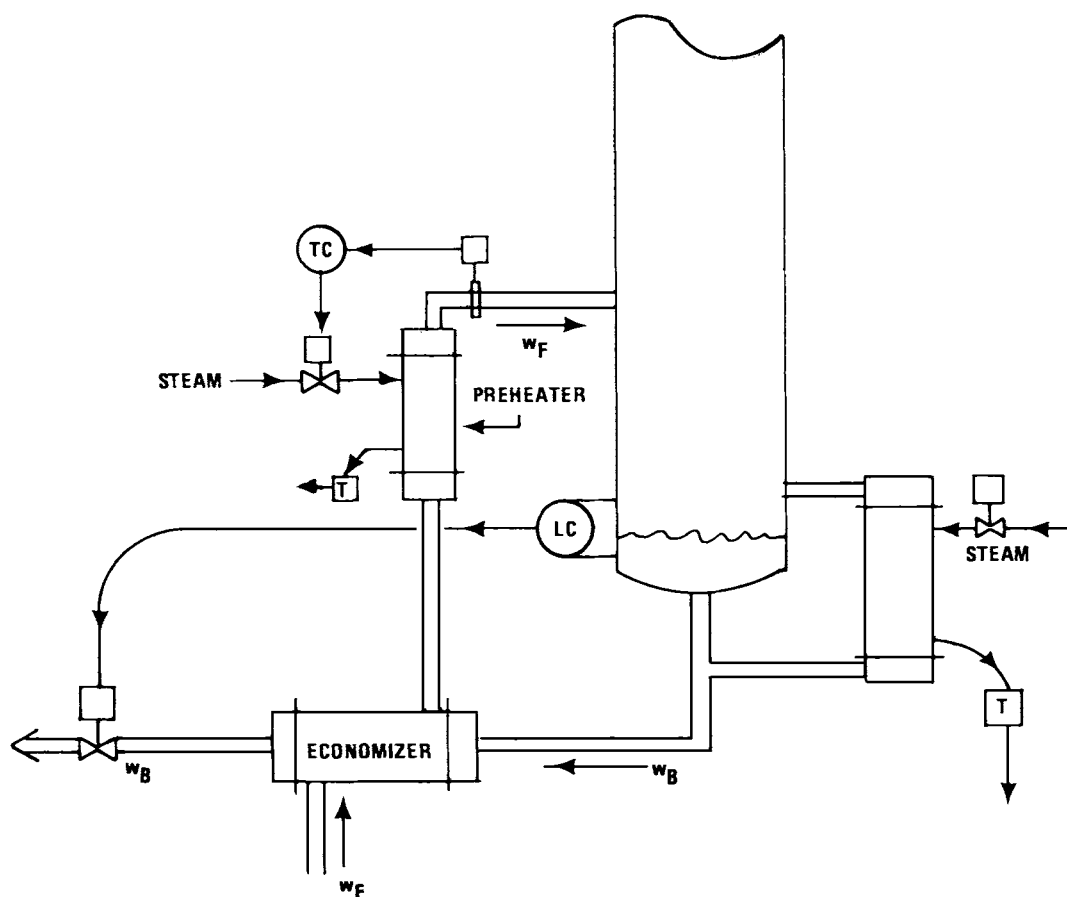


FIGURE 5.4
Column feed temperature control with economizer and preheater

enthalpy cannot be measured directly, it must be calculated as follows:

$$\psi_F = c_{pf} T_1 + \frac{w_B}{w_F} c_{pb} (T_{B1} - T_{B2}) + \frac{w_S}{w_F} \psi_S \quad (5.1)$$

where

ψ_F = enthalpy of feed entering column, pcu/lbm

c_{pf} = feed specific heat, pcu/lbm °C

T_{B1} = temperature, degrees Celsius, of bottom product entering economizer

T_{B2} = temperature, degrees Celsius, of bottom product leaving economizer

T_1 = feed initial temperature, degrees Kelvin (= °C + 273)

w_B = bottom-product flow rate, lbm/min

w_F = feed rate, lbm/min

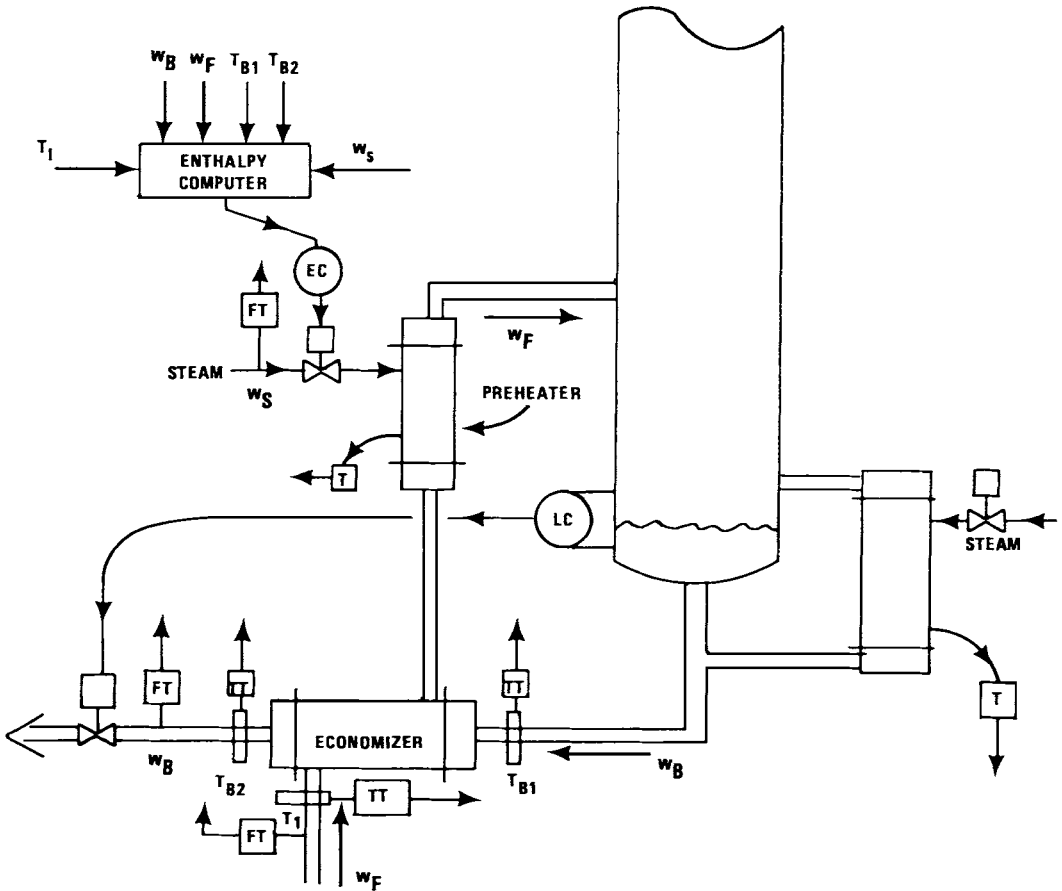


FIGURE 5.5
Column feed enthalpy control with economizer and preheater

c_{pb} = bottom-product specific heat, $\frac{\text{pcu}}{\text{lbm}} \text{ } ^\circ\text{C}$

w_s = steam flow rate to steam preheater, lbm/min

ψ_s = difference in enthalpy of steam entering preheater and condensate leaving it, pcu/lbm

The manipulation of feed enthalpy can be very important for some columns in utility optimization schemes and in maximizing column capacity or separation.

It is often possible to transfer heat into the feed stream using a lower temperature, less expensive heat source than must be used in the reboiler, since the base of the column is at the highest temperature in the column. Intermediate reboilers in the stripping section of the column are often used for the same reason.

Adding more heat to the feed reduces the amount of heat that must be added to the reboiler (but not in a one-to-one ratio) and increases the reflux ratio required to make the same separation. This increases the heat load on the condenser. If air or cooling water is used in the condenser, it is often economical to increase feed preheat. If refrigeration is required in the condenser, it is usually economical to minimize feed preheat.

The effect of feed preheat on column capacity can be significant. Feed preheat should be maximized when the column is reboiler limited or when it is limited by flooding below the feed tray. Feed preheat should be minimized when the column is condenser limited or is limited by flooding above the feed tray.

5.5 FEED TRAY LOCATION

For a given feed composition and enthalpy, there is an optimum feed-tray location that permits making the specified separation with the least energy consumption. It is also the tray that will permit maximum feed rate without causing the column to flood. As shown in Figure 5.6, a column should generally be equipped with a number of alternative feed trays to handle changes in operating conditions from those assumed for design.

The magnitude of the energy savings to be realized by changing feed tray location can be very significant in some systems (10–20 percent reduction in heat input), but in other columns the effects can be small. Each system must be examined to determine the strategy and the incentives for controlling to an optimum feed-tray location. Sometimes unexpected results occur. Luyben⁵ has shown that the optimum feed-tray location in some columns rises higher in the column as the feed becomes lighter (increase in more volatile component concentration), while in other columns exactly the opposite is true.

Many columns have more than one feed. Each feed stream should be introduced onto its optimum feed tray if energy consumption is to be minimized. These feeds should not be mixed together and introduced onto the same feed tray if their compositions differ.

5.6 FEED TANK SIZING

There are a number of possible criteria for tank sizing. They include the following.

Mixing. The mixing time constant of a well-mixed tank

$$\tau_M = \frac{\bar{V}}{\bar{Q}_F}$$

should be large compared with the reciprocal of the closed-loop natural frequencies of the downstream column-composition control loops. Here \bar{V} = average vessel holdup, ft^3 , and \bar{Q}_F = average feed rate, ft^3/min .

Flow Smoothing. The level control time constant τ_H should also be large compared with the reciprocal of the closed-loop natural frequencies of the downstream column-composition control loops.

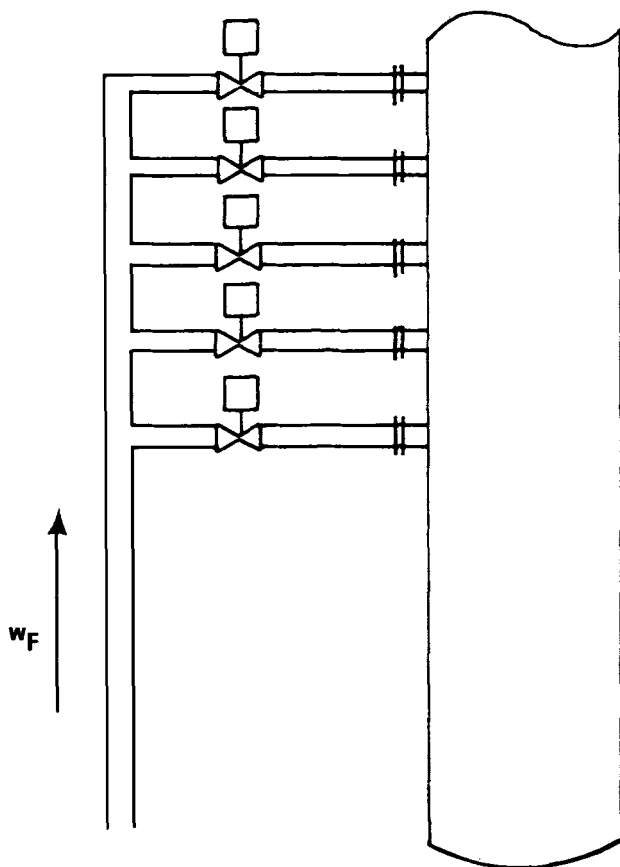


FIGURE 5.6
Column with multiple feed trays

The closed-loop natural frequencies of the composition control loops are not easy to calculate early in a project, but in general τ_M and τ_H should be at least 15–20 minutes. Tank (holdup) size is also a function of whether intermediate tanks will be used in a train or whether product will be transferred directly from one column base to another column. Generally speaking, composition control of a column will be easier (faster) if the reflux drum and column base are kept small. On the other hand, control of the succeeding process steps will be easier if the feed holdup is large. However, the use of intermediate tanks with supports, pump, electric switchgear, and instruments may increase investment substantially. Compromises, therefore, are frequently necessary. As a rough rule of thumb, if required τ_H or τ_M is much greater than 15–20 minutes, a separate surge tank should be used for column feed.

If the engineer is confronted with one of the following situations, then a separate surge tank for column feed is recommended:

1. Top- or bottom-product quality specification of the column to be fed is 0.1% (mole or weight) impurities or less.
2. Column to be fed is involved in a heat-recovery scheme where the vapor from one column furnishes part or all of the reboil heat for another column or columns.
3. Column to be fed has a sidestream drawoff that is a major fraction of the feed.
4. Column to be fed is expected to operate at or close to maximum capacity (very close to flooding) for a substantial fraction of on-stream time.
5. Columns in train have primitive controls (see Chapter 1).

If it is decided that a separate surge tank is needed, then, as a starting point, surge tank volume between the level taps should correspond to a $\tau_H \geq 1$ hour. For a PI controller with auto overrides and $K_{ch} = 0.25$ (see Chapter 16), $A \Delta H_T = 0.25 \tau_H \times (Q_o)_{\max}$ (vertical cylindrical tank assumed). The reset time for unity damping ratio and $\tau_H = 1$ hour would then be 4 hours. If this cannot be achieved with the proposed vendor's hardware, one should consider using a dedicated small computer as a level controller. Otherwise one must go to a proportional-only level controller, preferably with gain 2. Then $A \Delta H_T = 2 \tau_H \times (Q_o)_{\max}$. If item 4 is a significant consideration, proportional-only control is best since output flow changes are never larger than input flow changes.

The surge tank should also have good mixing; the recirculation rate should equal or exceed $10Q_{FS}$.

5.7 FEED SYSTEMS FOR DOUBLE-COLUMN SYSTEMS

Figures 5.7 to 5.9 illustrate several systems where two (or more) columns are operated in series or in parallel.

Figure 5.7 shows a simple split column where two columns are employed instead of one tall column, usually because of structural advantages. This scheme

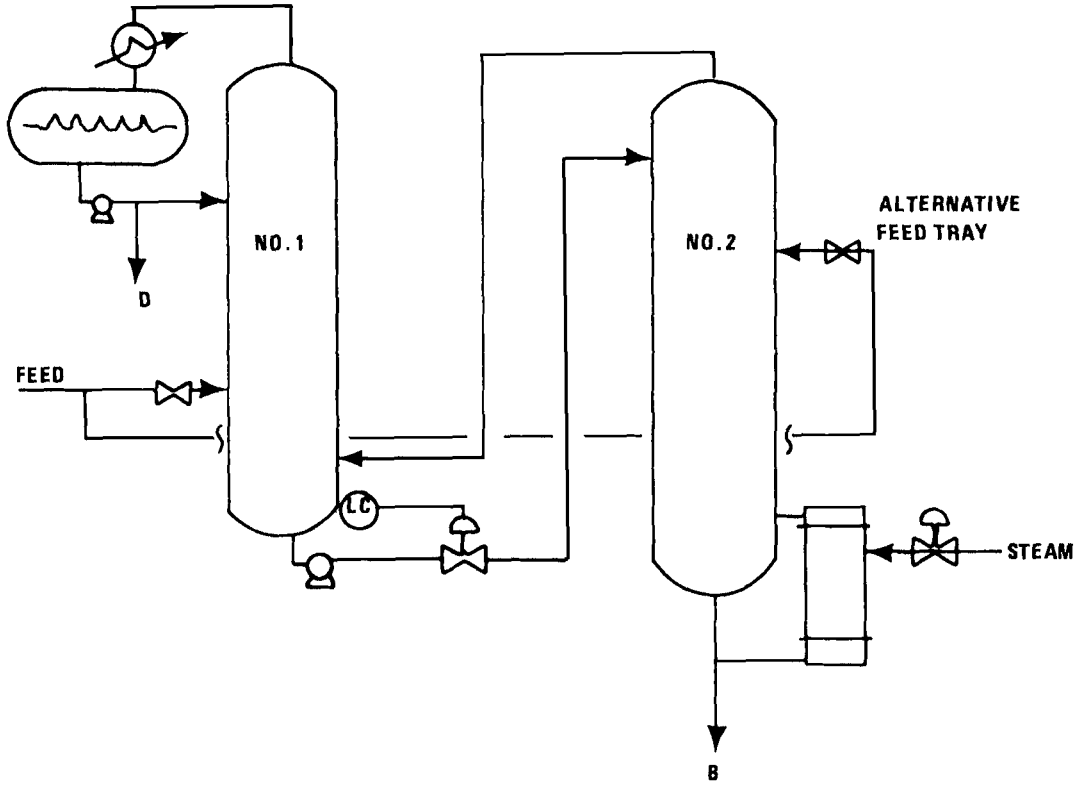


FIGURE 5.7
Feed system for a split column

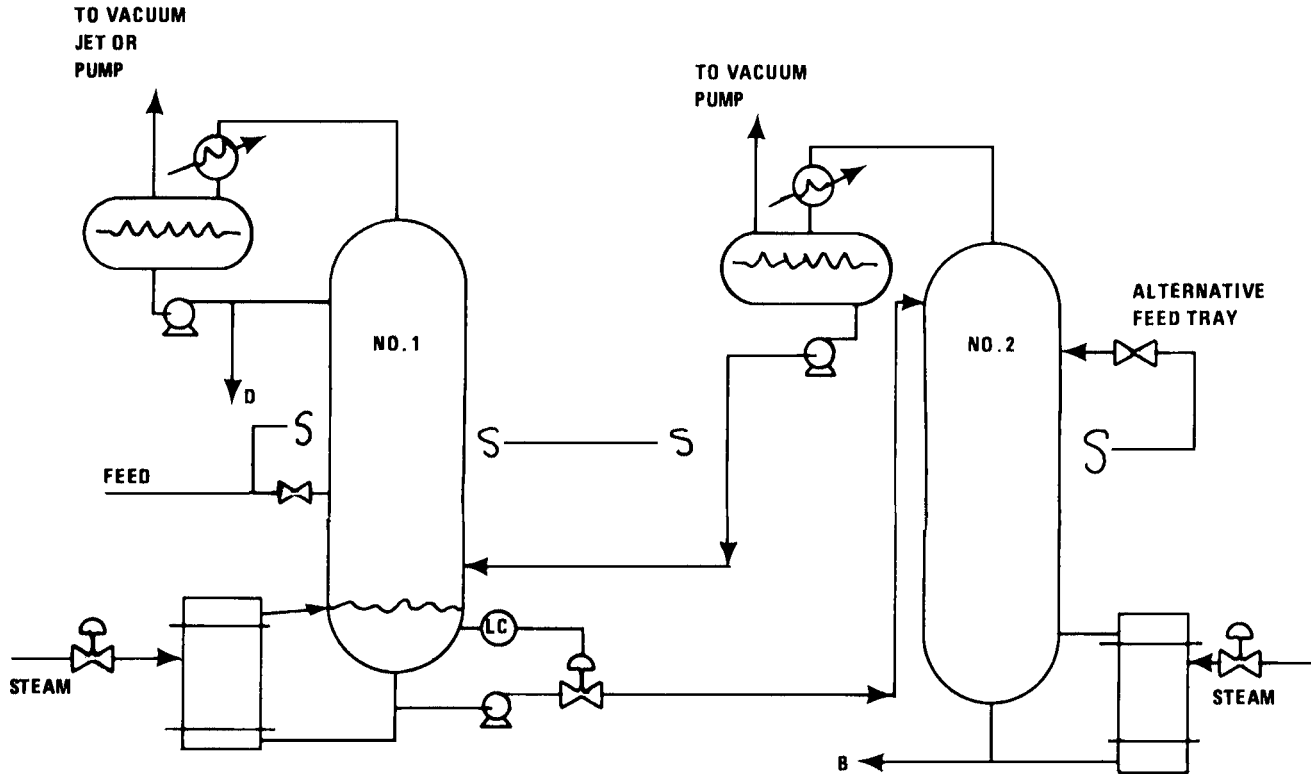
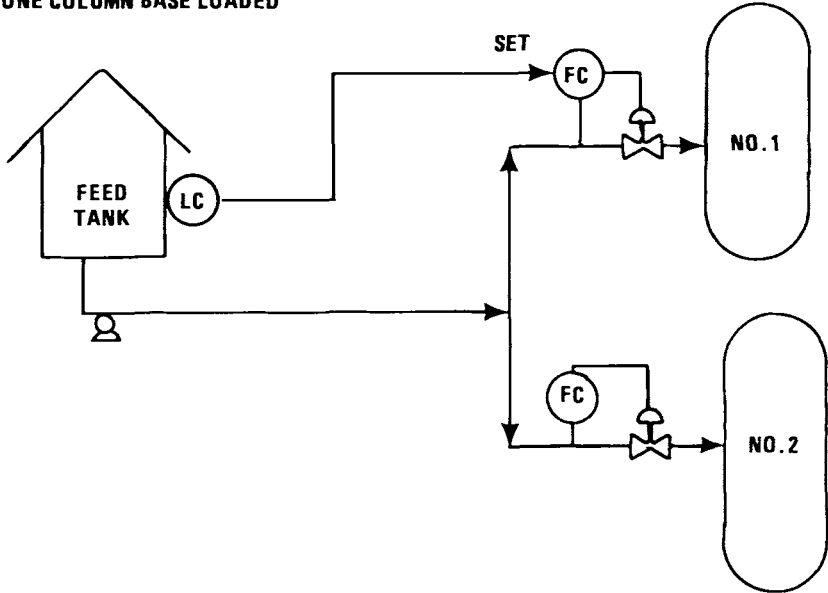


FIGURE 5.8
Feed system for split vacuum columns

A. ONE COLUMN BASE LOADED



B. BOTH COLUMN FEED RATES VARIED

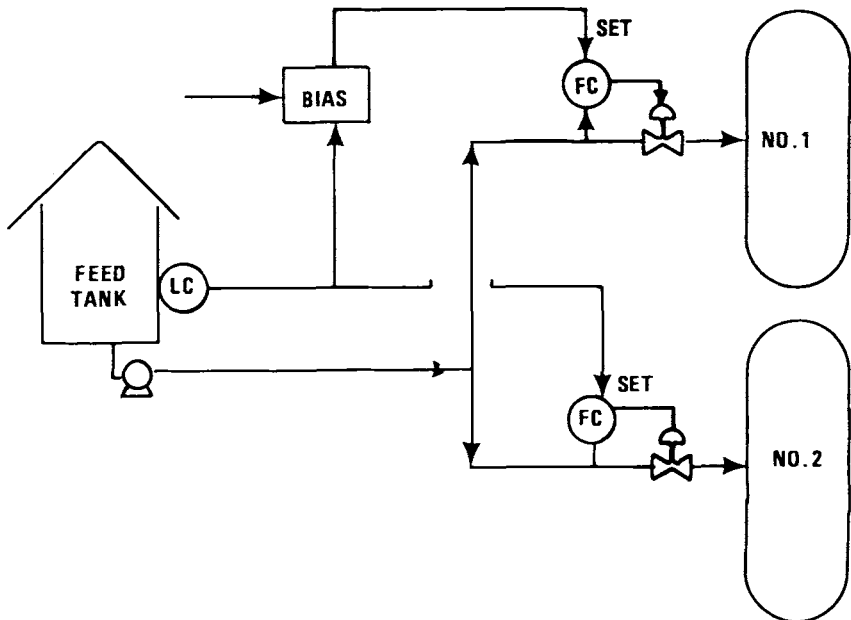


FIGURE 5.9
Feed systems for columns in parallel

is also sometimes used when revamping an existing process to utilize existing columns. All the vapor is generated in the base of the second column. The vapor from the top of the second column is fed into the base of the first column. Liquid from the base of the first column is fed onto the top tray of the second column. Feed can be introduced into either column, depending on the optimum feed-tray location.

Figure 5.8 shows a similar but distinctly different two-column system. It is sometimes used in vacuum systems if maximum base-temperature limitations and tray pressure drop prevent the separation from being achieved in one tall column. Each column has its own reboiler and condenser. The top of each column is maintained under vacuum. Liquid reflux for the second column is provided by the liquid from the base of the first column. Vapor from the top of the second column is condensed and pumped into the base of the first column. Feed can be fed to either column as dictated by composition. Note that this system uses about twice as much energy as a single column would require.

Figure 5.9 illustrates feed systems commonly used for two (or more) columns operating in parallel. The upper configuration, Figure 5.9A, is used when there is some advantage in keeping the feed rate to one of the columns constant. The other column then takes the swings in feed rate. If the two columns are of unequal capacity, the smaller column is normally the one that would be base loaded since swings in feed rate to the larger column are smaller on a percentage basis.

The lower configuration, Figure 5.9B, permits both feed rates to vary. A bias signal is usually needed to balance the feed rates between the columns.

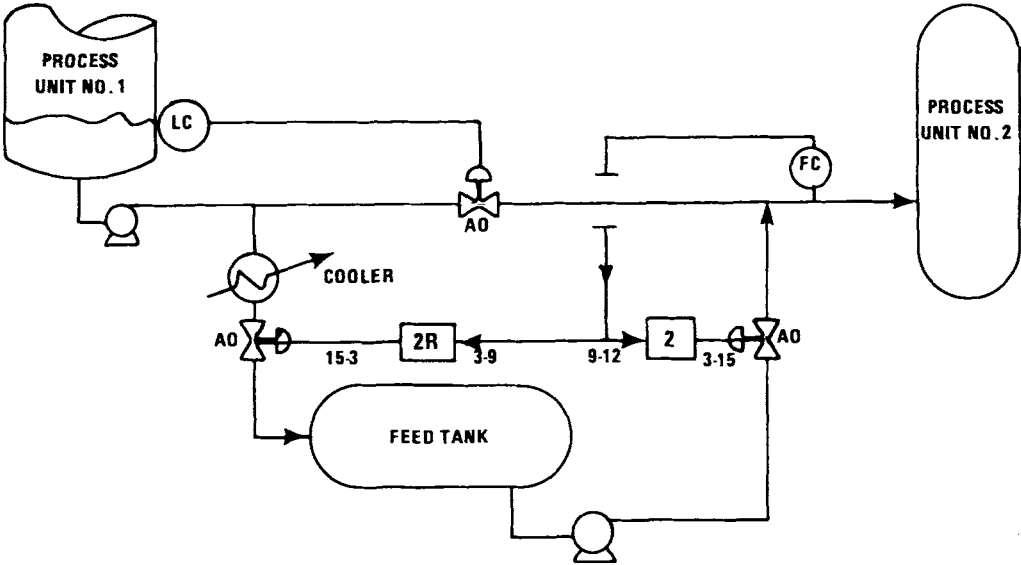
5.8 FEEDS WITH MAKEUP/PURGE TO TANKAGE

Figure 5.10 shows a system sometimes encountered when a large feed tank is used to balance load and demand at some point in the process. The tank is connected in series with the main process flow. Most of the flow from Process Unit No. 1 goes directly to Process Unit No. 2. The level in Process Unit No. 1 must be controlled; at the same time, it is desired to fix the flow rate into Process Unit No. 2.

It is usually undesirable to put everything through the feed tank, since this typically operates at a lower temperature than Process Unit No. 1. Cooling all the process stream and then feeding it to Process Unit No. 2 normally increases energy consumption in process Unit No. 2 significantly.

The control system shown in Figure 5.10A is a simple technique for minimizing energy consumption. The two valves on the makeup and purge lines from and to the feed tank are split ranged so that both valves cannot be open at the same time. This guarantees that a minimum amount of material is going to or coming from the tank.

A. SINGLE SYSTEM



B. MORE COMPLEX SYSTEM PROVIDING COMPLETE ISOLATION OF LOOPS

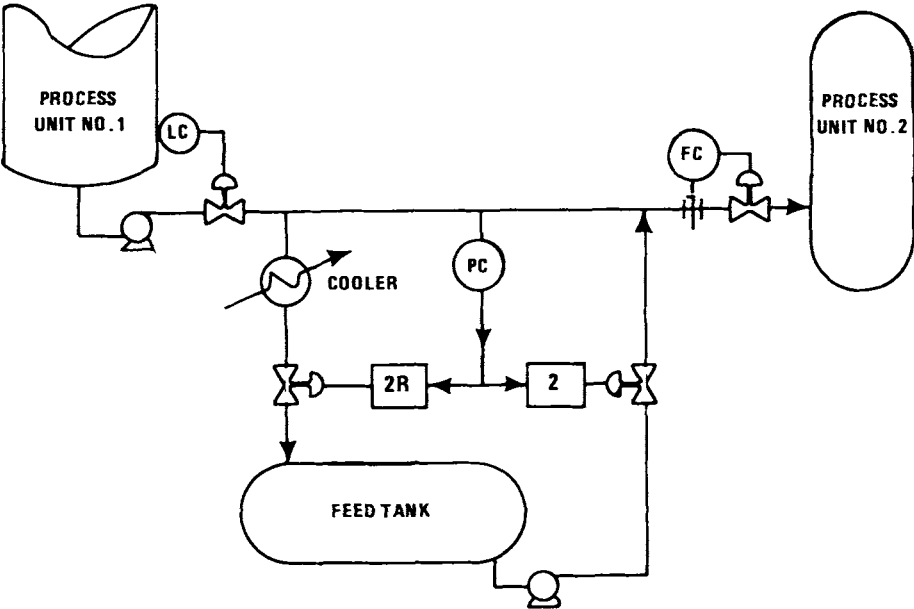


FIGURE 5.10
Makeup/purge feed systems

The system has some dynamic problems. The level loop is affected by the flow loop. This problem can be eliminated by the more sophisticated and costly system shown in Figure 5.10B. A pressure controller on the “header” is used to balance supply and demand flow.

The system in Figure 5.10A also requires overrides:

1. To prevent the level in Process Unit No. 2 from getting too low (a low level pinches the valve in the line to the feed tank).
2. To prevent too much material from entering Process Unit No. 2 (if the flow gets too high, the valve in the direct line connecting to two units is pinched).

5.9 FEED SYSTEMS IN SEQUENCES OF COLUMNS WITH AND WITHOUT RECYCLES

Trains of distillation columns are often encountered. Material flows through these sequences of columns, and small fluctuations in flow rates in upstream columns can be amplified as they work their way down through the train if two or more level controls in sequence have unenhanced PI controllers and have the same τ_H and τ_R . Although this is uncommon, it may happen. The engineer, therefore, should check all level controls in a sequence of process equipment. If necessary, some of the tank sizes or controller parameters may have to be changed.

It has been shown by one of the authors, in an unpublished study, that if one uses PI controllers enhanced with auto-overrides, this problem is greatly alleviated.

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6 Level Control and Feedforward Options

6.1 INTRODUCTION

As indicated in Chapter 1, it is most convenient, when starting the design of controls for a new or modernized plant, first to lay out all of the material-balance controls. These are mostly liquid level controls. It was also indicated that feedforward compensation could be used to supplement feedback composition controls to achieve more constant compositions. In the absence of feedback composition controls—usually because adequate composition measurements are lacking—feedforward compensation is almost mandatory.

In this chapter we will consider only feedforward compensation for production rate changes. These are usually larger and more rapid than composition or thermal changes. The most common compensation technique is that of ratio controls, either open loop via a multiplier or closed loop via a divider. It has been shown^{1,2} that for single-loop (no cascade) control, either technique is satisfactory, provided the proper installed valve flow characteristic is used. For cascade loops, however, where the flow-ratio control is the secondary or slave loop, there is a serious variation of control loop gain from low to high production rates.² Consequently, as discussed in Chapters 12 and 20, it is usually advisable to use an impulse feedforward technique unless the wild flow rate changes less than 2:1. For simplicity, however, we will use ratio controls in most of the illustrations in this chapter.

We will begin with combinations of level control and feedforward compensation for applications where material-balance control is in the direction opposite to flow. Then we will consider schemes in which material-balance control is in the direction of flow. Unfavorable schemes—those that are hard to design or to make work—will be pointed out; their use should be avoided unless no suitable option is available.

Only conventional columns with top and bottom drawoffs will be considered here. Columns with sidestream drawoffs are discussed in Chapter 7. Note that

if reflux drum level control sets reflux, or if base level sets steam, level control must be as “tight” as possible for best composition control.

6.2 MATERIAL-BALANCE CONTROL IN DIRECTION OPPOSITE TO FLOW

This approach implies that either the distillate or the bottom product is the demand flow and that the column must be operated at a rate to satisfy that demand. We can now visualize several possible designs.

Bottom-Product Demand; Base-Level Control Via Feed

Bottom-product demand requires, in most cases, that base level be controlled by feed rate. Then there will be two options for reflux drum level control.

Overhead Level Control Via Top Product

There has been considerable controversy in the literature about whether to have reflux drum level control via top product or via reflux flow.

Our own studies fail to show any overwhelming advantages one way or the other. On the average, however, we find top composition control somewhat more straightforward when reflux is manipulated by a ratio controller or by a direct composition controller as shown by Figure 6.1.

Since the bottom-product flow is the demand flow, steam and reflux are ratioed to it. For all ratio loops, appropriate dynamic feedforward compensators should be provided.

Since there is a dead time involved—the time for liquid to flow from the feed tray to the column base—the base level controller settings should be determined by the method of Chapter 16, Section 6. This section also gives a complete design, including overrides for bottom product and steam flows. The reflux drum level controller settings (if a PI controller is used) should be determined by the method of Chapter 16, Section 2.

Overhead composition may be controlled by trimming the reflux/bottom-product ratio. Base composition may be controlled by trimming the steam/bottom-product ratio.

Reflux Drum Level Control Via Reflux

In this case, as shown in Figure 6.2, we have steam/bottom-product and distillate/bottom-product ratio controls.

In calculating reflux drum level controller settings, whether proportional only or PI, one should take care to account for reflux subcooling (see Chapter 16, Section 4). Overhead composition may be controlled by trimming the distillate/bottom-product ratio with perhaps a feedforward compensator connected into the overhead level control loop. Base composition may be controlled by trimming the steam/bottom-product ratio control.

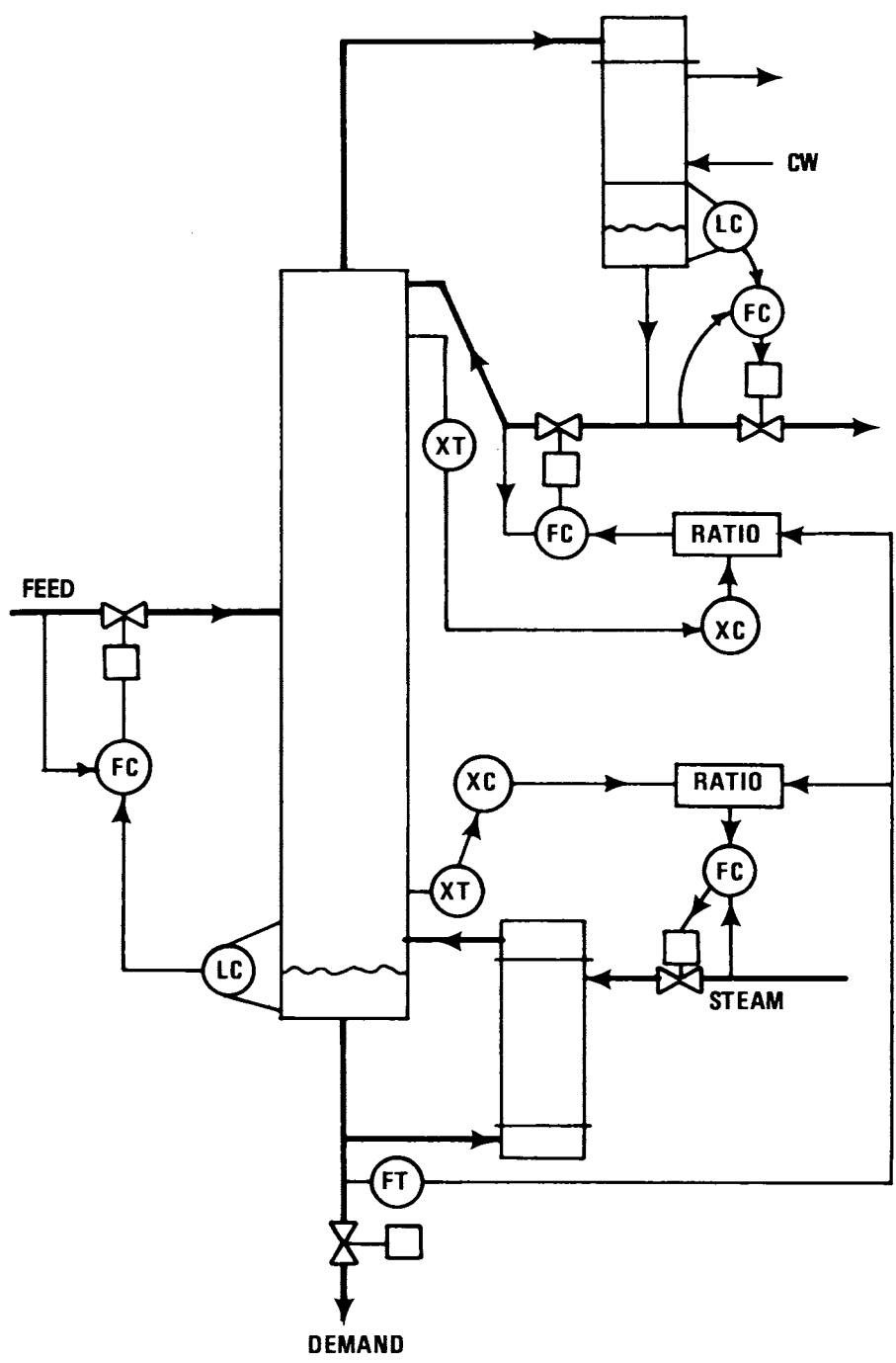


FIGURE 6.1
Bottom product demand, overhead level control via top product, base level via feed

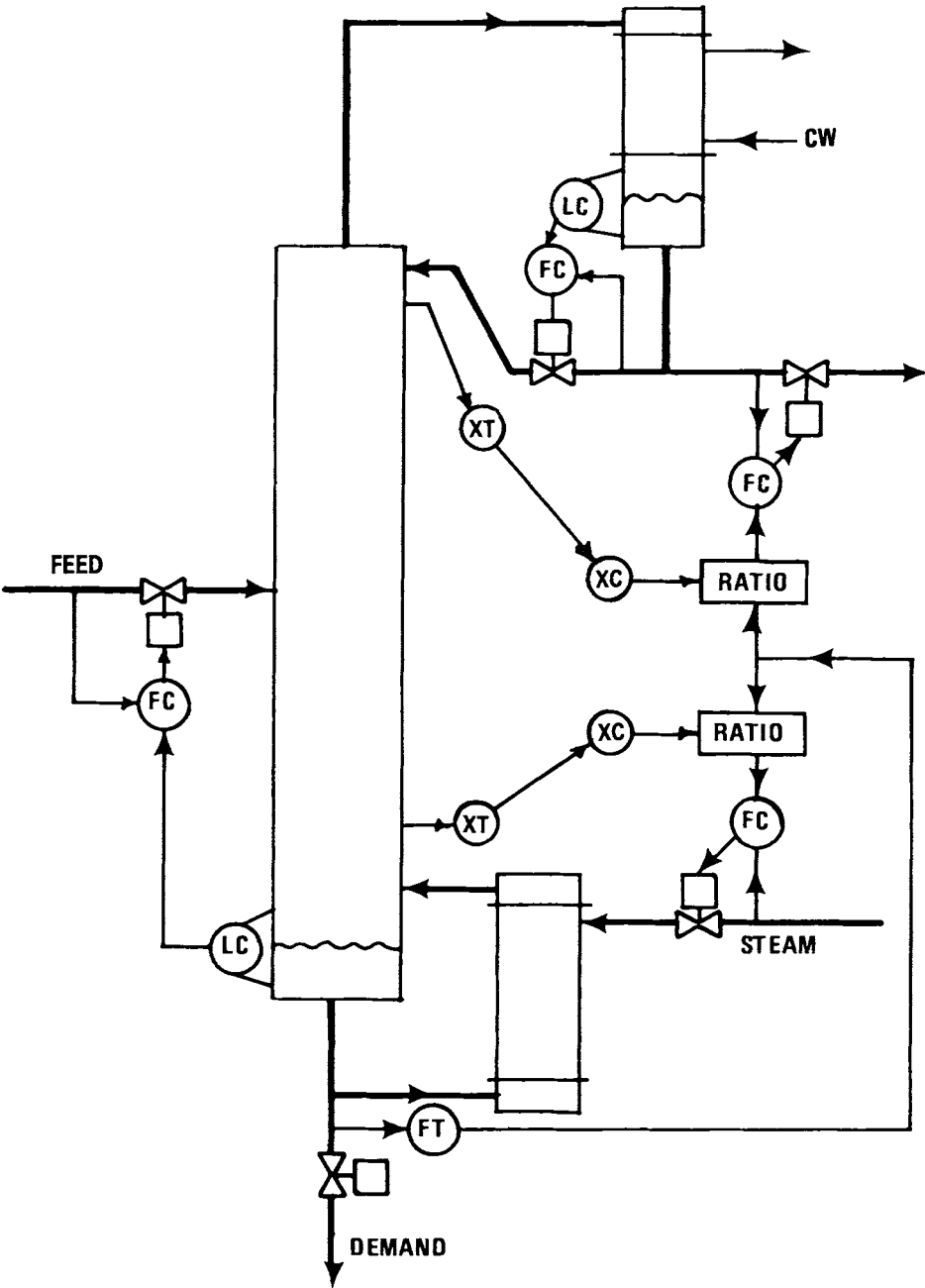


FIGURE 6.2
Bottom product demand, overhead level control via reflux, base level control via feed

Reflux Drum Level Control Via Boilup

As shown in Figure 6.3, this case requires reflux/bottom-product and distillate/bottom-product ratio controls.

Determining settings for the reflux drum level controller is, in this case, difficult unless a large reflux drum holdup is available. Preferably one should make $\tau_H \geq 5$ minutes; level controller tuning will require a dynamic analysis of overall column material balance such as discussed in Chapter 14. If steam flow is metered by an orifice, it should be linearized with a square root extractor.

Base level controller settings may also be determined by the method of Chapter 16, Section 6. Overhead composition control may be accomplished by trimming the top-product/bottom-product ratio. Base composition may be controlled by trimming the reflux/bottom-product ratio. Because of interactions between composition controls, this column control scheme is not desirable, although probably not impossible.

Distillate (Top-Product) Demand; Base Level Control Via Feed**Reflux Drum Level Control Via Reflux**

This case requires, as shown by Figure 6.4, bottom-product/top-product and steam/top-product ratio controls.

The calculation of reflux drum level controller settings follows the method of Chapter 16, Section 4; the calculation of base level controller settings follows that of Chapter 16, Section 6.

Since reflux drum holdups are usually small compared with base holdups, a buffer tank in the top product line (*not* in the reflux line) is highly recommended. Top product composition may be controlled by trimming the steam/distillate ratio; bottom composition may be controlled by trimming the bottom-product/distillate ratio.

Reflux Drum Level Control Via Boilup

As shown in Figure 6.5, this scheme requires reflux/distillate and bottom-product/distillate ratio controls. A buffer tank in the top-product line is recommended.

As indicated earlier, reflux drum level control via boilup is difficult unless a lot of holdup is available. Overhead composition may be controlled by trimming the reflux/distillate ratio; base composition may be controlled by trimming the bottom-product/distillate ratio. Note that if either distillate or bottom product (or side product) is a demand flow, either reflux drum or base level control must manipulate feed rate.

6.3 MATERIAL-BALANCE CONTROL IN DIRECTION OF FLOW

Practically speaking, material-balance control in the direction of flow means that the column must take whatever feed is supplied. This is, however, subject

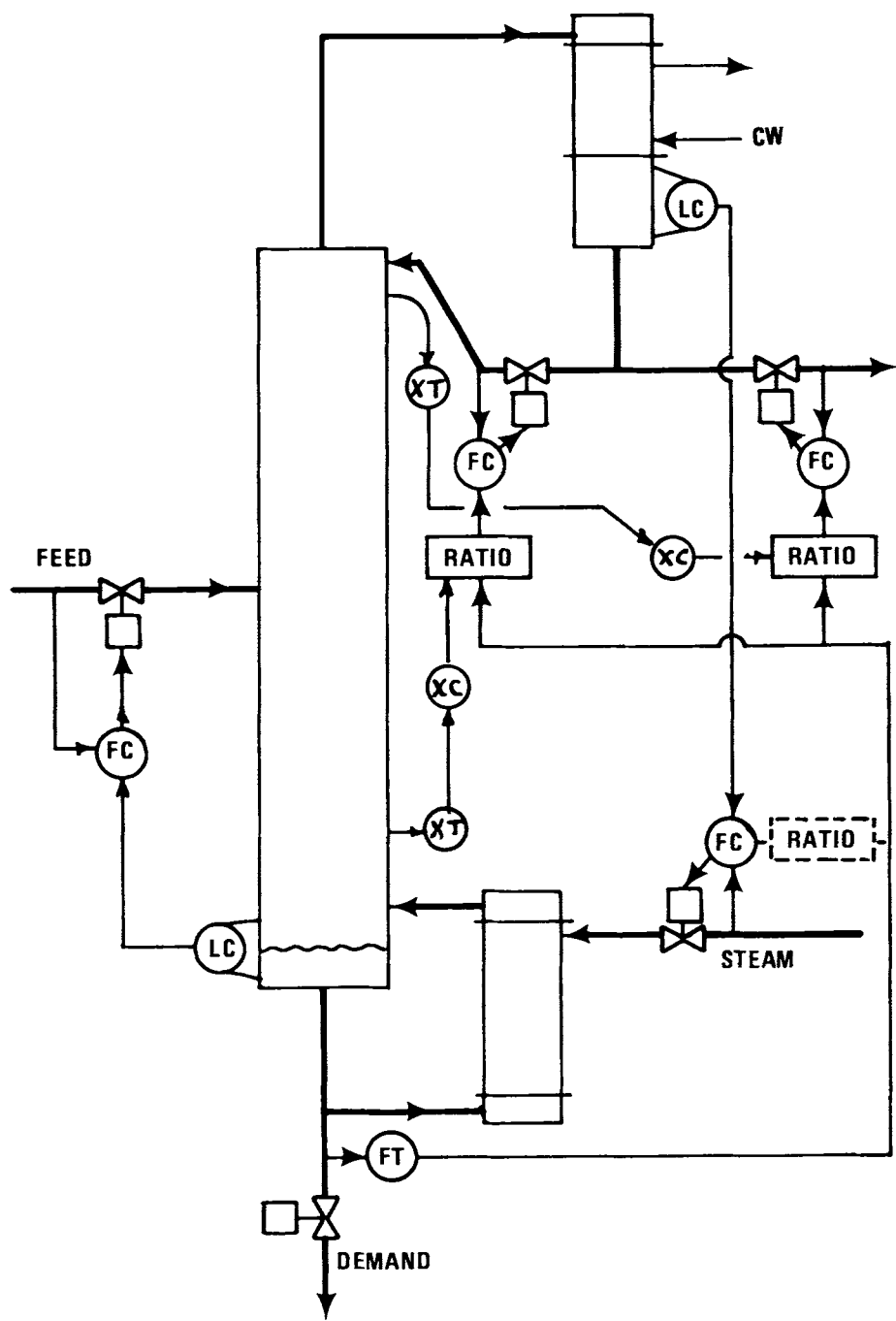


FIGURE 6.3
Bottom product demand, overhead level control via boil up, base level control via feed

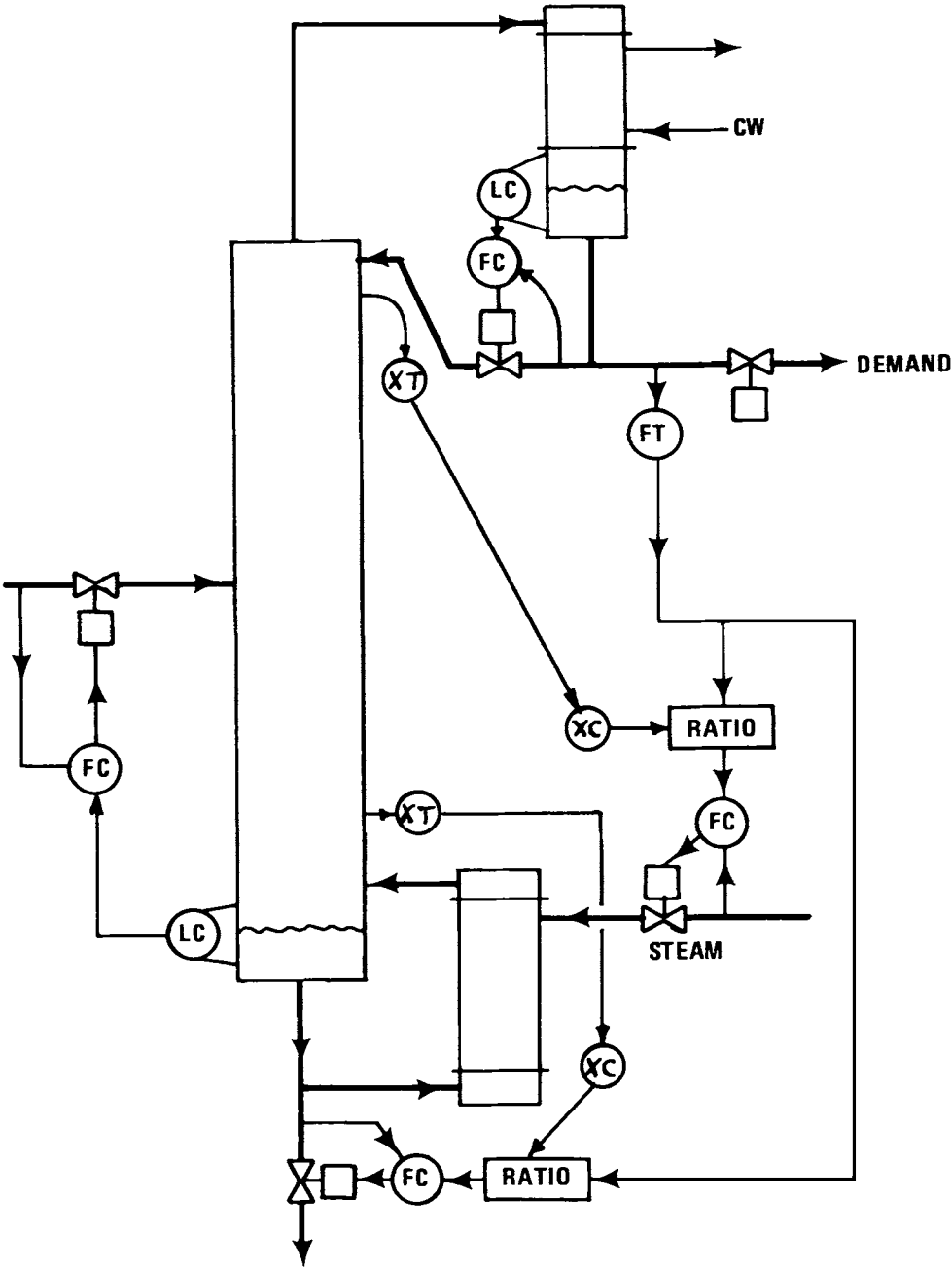


FIGURE 6.4
Distillate demand, reflux drum level control via reflux, base level control via feed

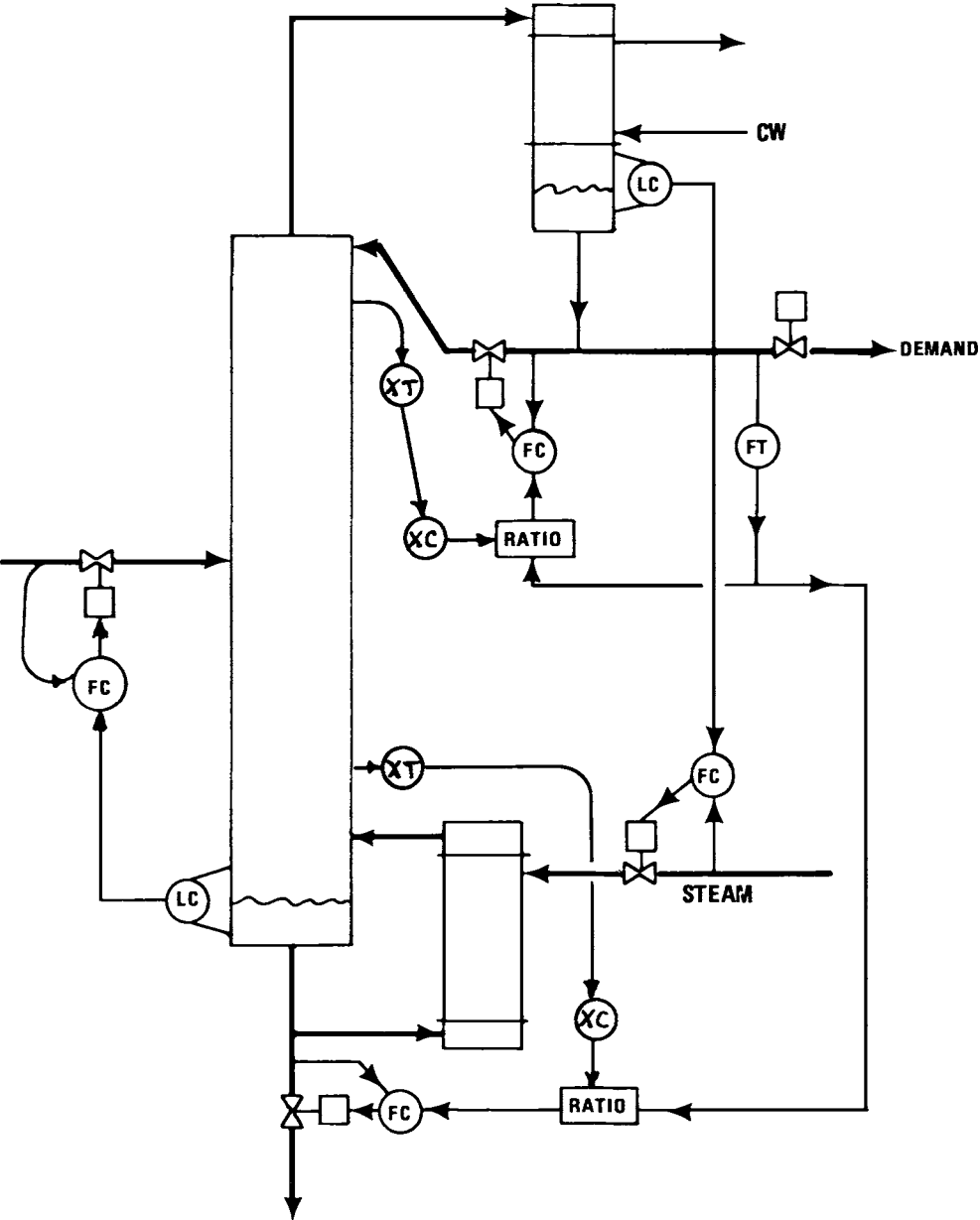


FIGURE 6.5
Distillate demand, reflux drum level control via reflux, base level control via feed

to the restriction that column overrides, particularly high base level, can restrict feed flow.

Reflux Drum Level Control Via Distillate; Base Level Control Via Bottom Product (Figure 6.6)

This is one of the most commonly encountered column-control schemes. Both level control systems may be calculated by the method of Chapter 16.

Reflux/feed and steam/feed ratio controls should be provided. Top composition may be controlled by trimming the reflux/feed ratio; bottom composition by trimming the steam/feed ratio.

Reflux Drum Level Control Via Reflux; Base Level Control Via Bottom Product (Figure 6.7)

This is also a commonly encountered control scheme. Both level controls may be calculated by the method of Chapter 16; care should be taken to include the subcooled reflux effect on overhead level control.

Distillate/feed and steam/feed ratio controls should be provided. Top composition may be controlled by trimming the distillate/feed ratio and with perhaps additional feedforward to the reflux drum level controller. Base composition may be controlled by trimming the steam/feed ratio.

In many cases this scheme will require a buffer tank in the top-product line to the next piece of process equipment, unless the top product is going directly to storage.

Reflux Drum Level Control Via Distillate; Base Level Control Via Boilup (Figure 6.8)

Overhead level control may be calculated simply by the method of Chapter 16, Section 3, but base level control by boilup is very difficult. It is normally used only when the average bottom-product flow is very small. The characteristic time constant τ_H should be at least 15 minutes and other design factors should be as indicated in Chapter 16, Section 7. In most cases base level control by boilup requires a dynamic analysis, and perhaps supplementary plant tests. If steam flow is measured with an orifice, a square root extractor should be used.

Reflux/feed and bottom-product/feed ratio controls should be provided. Top composition may be controlled by trimming the reflux/feed ratio while base composition may be controlled by adjusting the bottom-product/feed ratio.

Reflux Drum Level Control Via Distillate; Base Level Control Via Boilup; Reflux Ratioed to Top Product

It is sometimes recommended that reflux be ratioed to distillate to diminish interactions between top and base feedback composition controls.³ If, however, reflux drum level is controlled by manipulating top product, and base level is controlled by manipulating boilup (see Figure 6.9), there will be a dynamics

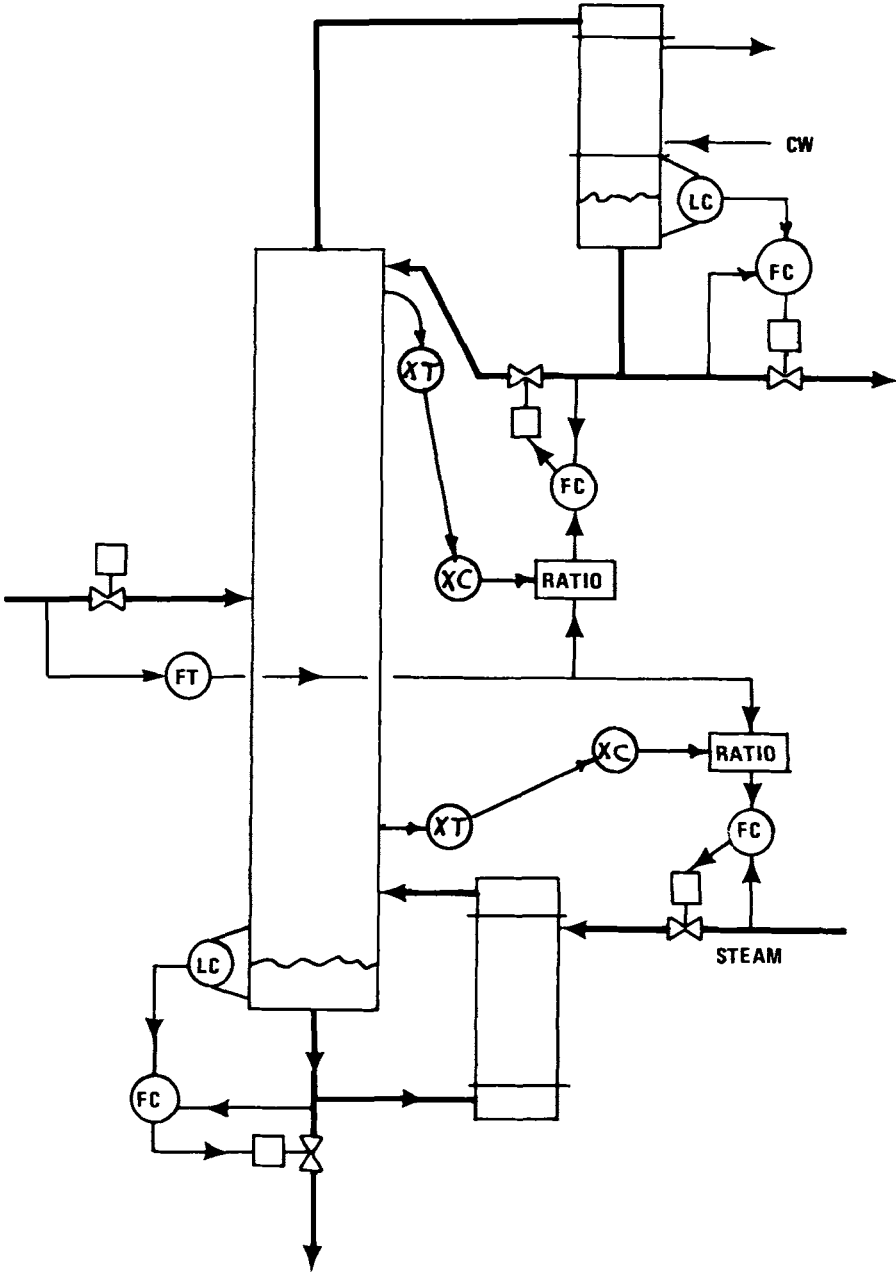


FIGURE 6.6
Material balance control in direction of flow, reflux drum level control via distillate, base level control via bottom product

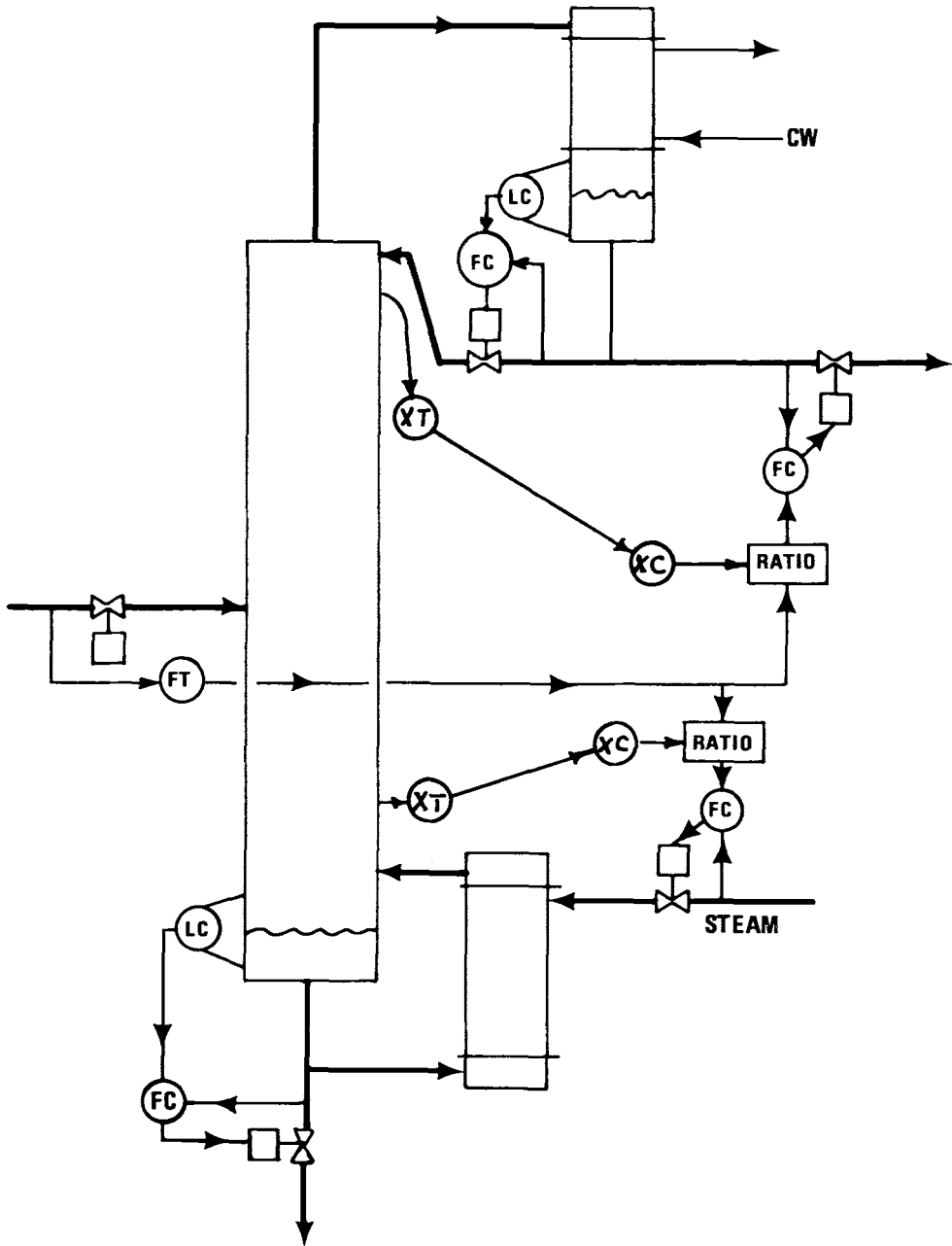


FIGURE 6.7
Material balance control in direction of flow, reflux drum level control via reflux, base level control via bottom product

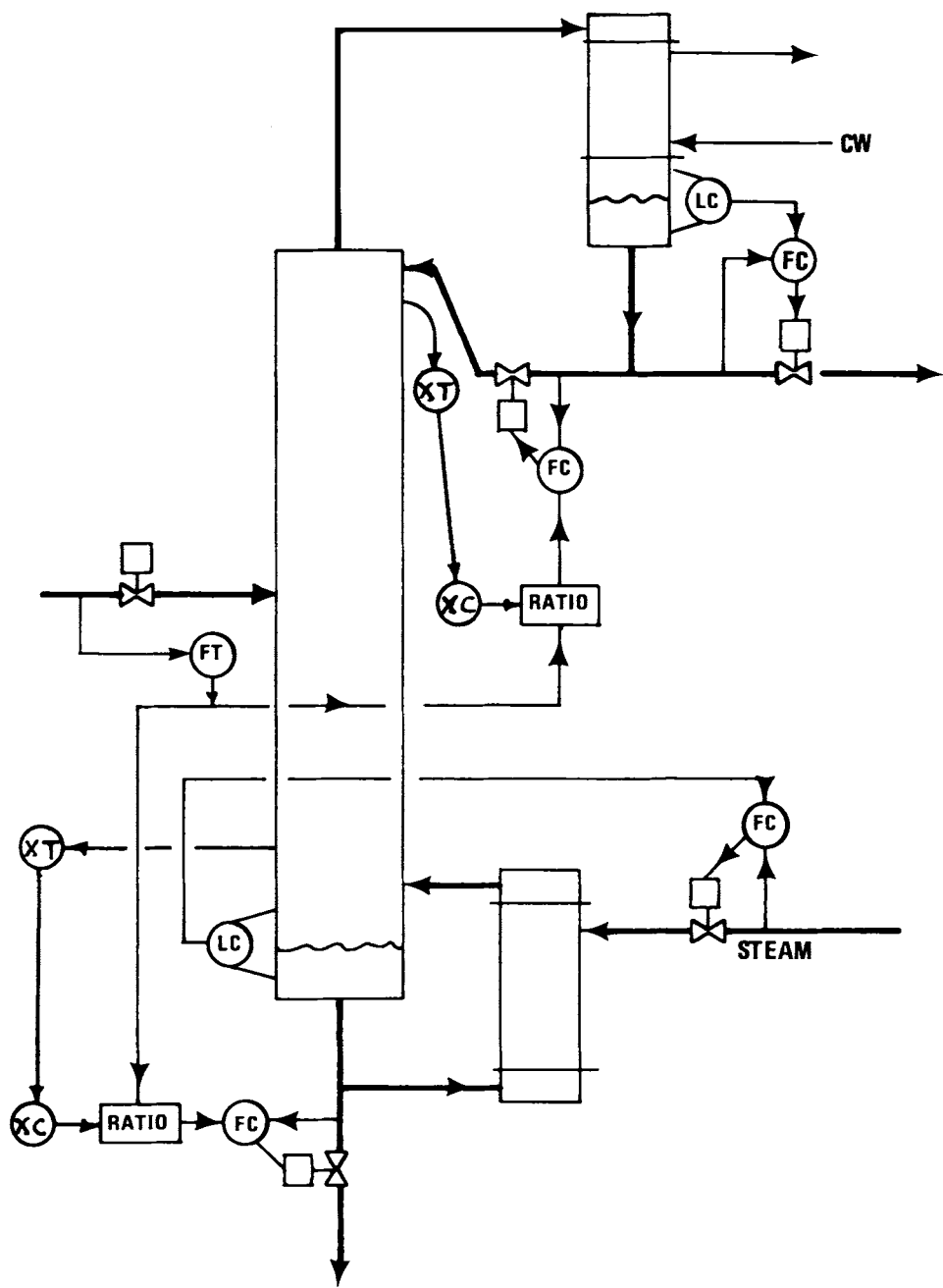


FIGURE 6.8
Material balance control in direction of flow, reflux drum level control via distillate, base level control via boilup

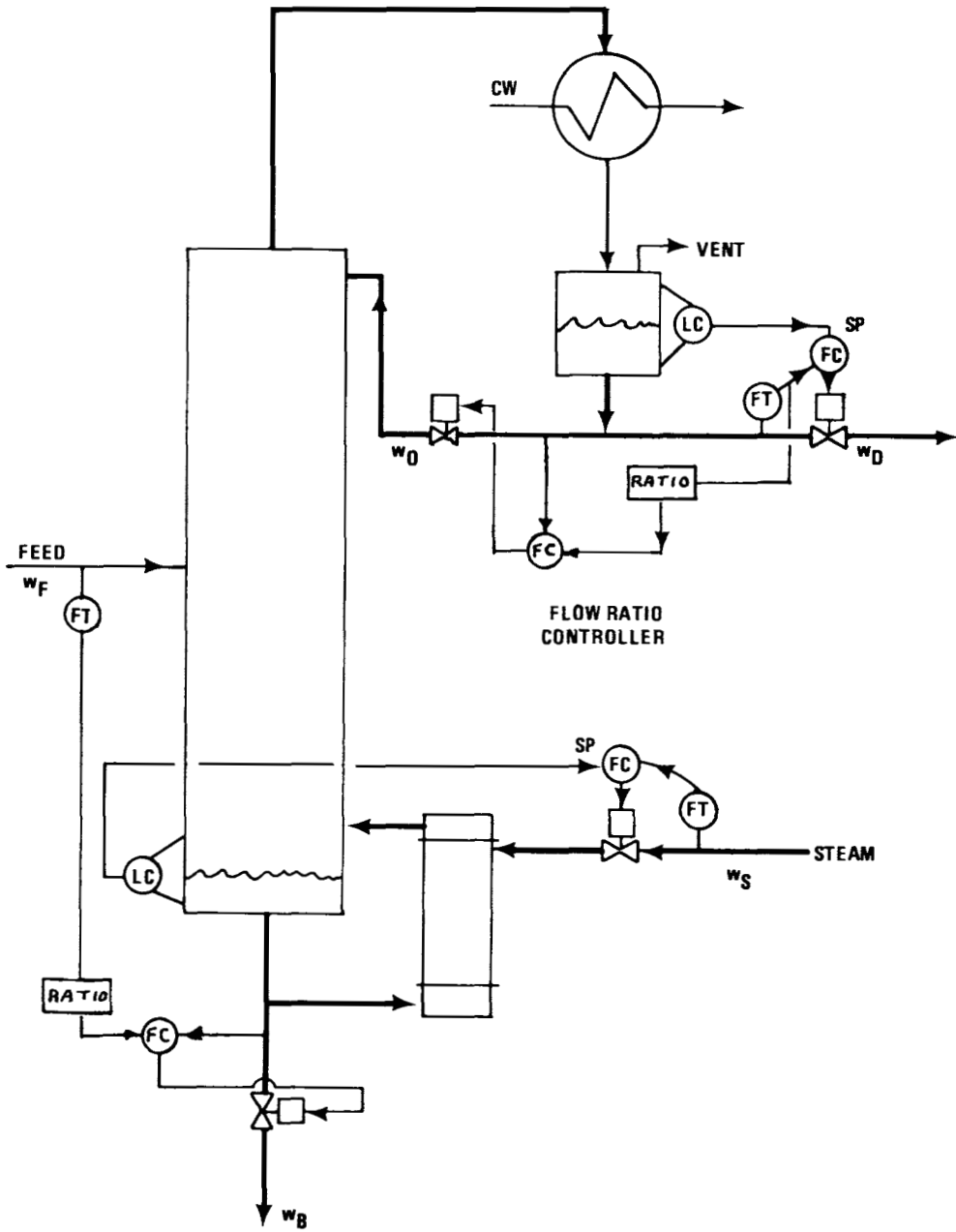


FIGURE 6.9
Like figure 6.8 but with reflux ratioed to distillate

problem since reflux is a function of boilup. This is in addition to other problems of controlling base level by boilup (see Chapters 4 and 16).

6.4 UNFAVORABLE CONTROL SCHEMES

Perhaps the most common system that has given trouble is base level control via steam. This is particularly true if a thermosyphon reboiler is employed or if the column has valve trays, or both. It is the result of “inverse response” (see Chapter 13). At low boilup rates, sieve trays give the same trouble. To minimize difficulties the design recommendations of Chapter 16, Section 7, should be followed. One of the authors has shown, in an unpublished study, that an “inverse response compensator” can be designed and implemented on a computer or with some microprocessor controls.

In a recent project we had considerable difficulty with base level control via feed accompanied by steam/feed ratio control. In effect the level controller adjusts both feed and steam. Since bottom product was the demand flow, this was corrected by converting the steam/feed ratio control to steam/bottom-product ratio control.

High base-level overrides on steam have also given trouble for the reasons mentioned. To a considerable extent, however, these problems can be mitigated by using proportional-reset instead of proportional-only overrides (Chapter 9). A high base-level override on steam should not be used on the same column with either reflux drum level control on reflux or a high reflux drum level override on reflux.

6.5 UNREASONABLE CONTROL SCHEMES

The control schemes in this section have one common fault—the inability to control the overall column material balance. Changes in the demand flow rate are not accompanied by appropriate changes in the other flows.

One of the worst combinations of controls is reflux drum level via reflux and base level via boilup. This configuration produces a positive feedback loop that either will shut the column down or open the steam and reflux valves wide, neither of which is a desirable situation. Operator intervention is eventually required. We have seen it used on columns with very large base holdups—8 to 24 hours.

If the bottom product is the demand stream, then base level control via boilup and reflux drum level control via distillate will not control the column properly. An increase in bottom product will cause a decrease in distillate, with no change in feed rate.

If the distillate is the demand stream, then base level control via bottom product and overhead level control via boilup will not control the column properly. This scheme is analogous to the preceding scheme.

If distillate, bottom product, or side product is a demand flow, one of the level controllers *must* manipulate feed rate.

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7

Control of Sidestream Drawoff Columns

7.1 INTRODUCTION

to reduce investment and energy consumption, column designers sometimes tackle the separation of multicomponent mixtures with one sidestream drawoff column instead of using two or more conventional columns, each with two product drawoffs. The most extreme example of this philosophy is the crude column or crude still of oil refineries. These columns not only are very large in diameter, commonly 25–40 feet, but also have a large number of product drawoffs and auxiliary heat exchangers. For the chemical industry, it is more common to see columns with a single side draw. In this chapter we will concern ourselves with this type of column. Even the simplest side drawoff column is usually much more difficult to control than conventional two-product columns. Further, such columns generally have much less flexibility and less turndown capability.

Whether to take side draw as a vapor or as a liquid is usually decided as follows¹:

1. If side draw is taken *below* the feed tray, it is usually taken as a vapor. This is done for the purpose of minimizing high boilers in the sidestream.
2. If side draw is taken from a point *above* the feed tray, it is usually taken as a liquid. The objective here is to minimize low boilers in the sidestream.

7.2 SIDE-DRAW COLUMNS WITH LARGE SIDESTREAMS

Sidestream drawoff columns sometimes have the task of removing small amounts of both low boilers and high boilers from a large intermediate boiler. Solvent-recovery systems in plants manufacturing plastics or synthetic textile fibers often utilize such columns.

Figure 7.1 shows a typical control scheme for a column with a vapor side draw. Since the top and bottom products are small, base level is controlled by throttling side draw. When the side draw is a vapor, column base holdup should be generously sized. Changing the side draw changes vapor rate to the top of the column; this changes the rate of condensation, which finally changes reflux back down the column. The overhead level control loop that manipulates reflux flow is nested within the base level control loop. Top draw (distillate), bottom draw (bottom product), and steam flow are shown ratioed to feed. If this is not done, composition control will be poor, column capacity will be limited, and steam will be wasted.

A practical problem is that of maintaining adequate reflux down the column. As will be shown in Chapter 9, this requires overrides or limiters for the side draw; otherwise the base level controller on occasion may cause so much side draw to be taken that reflux temporarily will be inadequate.

An alternative arrangement is shown in Figure 7.2. Here the side draw is a liquid. The downcomer, instead of emptying into the tray below, empties into a small surge vessel. Base level is controlled via side draw while surge-tank level control sets reflux back down the column. This scheme has been used successfully on a nonequal molal overflow distillation. By measuring the reflux flow being returned from the surge tank, we can use an override on the side draw to ensure adequate reflux.

As an alternative one might control surge vessel level by side draw and base level by reflux from the surge vessel.

7.3 SIDE-DRAW COLUMNS WITH SMALL SIDESTREAMS

As mentioned in Section 1.9, a distillation column sometimes tends to collect intermediate boilers, compounds that are heavier than the light key but lighter than the heavy key. In this case a small sidestream is required. As shown in Figure 7.3, the column-control scheme is very similar to that of a conventional two-product column with the addition of sidestream controls. Ideally the side draw should be ratioed to the feed, but in practice, if it is very small, simple flow control is sometimes used.

7.4 COMPOSITION CONTROL OF SIDE-DRAW COLUMNS

As indicated in Chapter 1, exact composition control of product streams requires as many streams as components, and as many manipulated variables as components. Let us consider how to provide composition control of the system in Figure 7.1.

Here top- and bottom-product flows (or their ratios to feed) are available for composition control, but the side draw is needed for column material-balance control. For this column, however, composition of the sidestream is of primary importance. Let us assume that the feed consists of lumped low

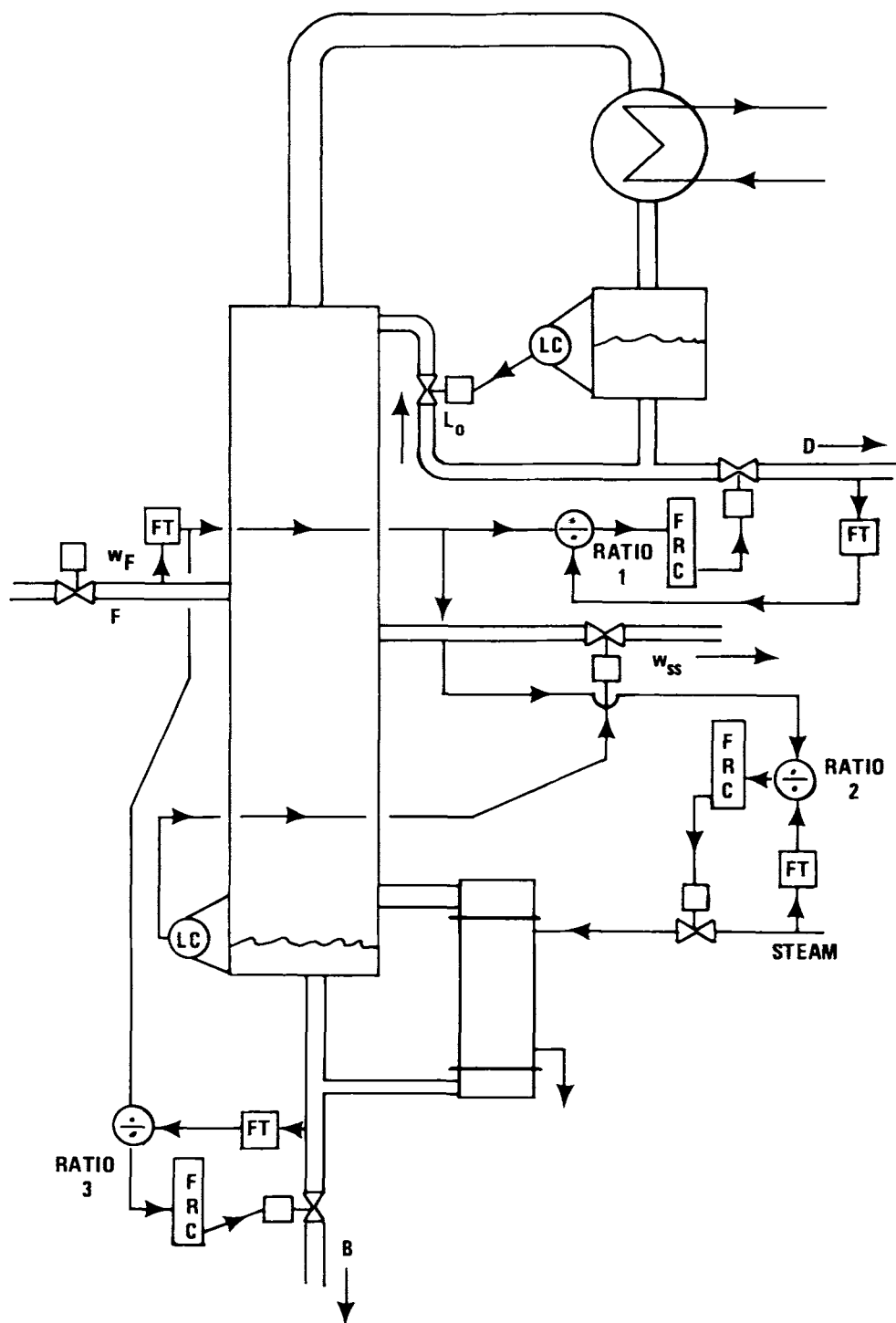


FIGURE 7.1
Basic control scheme for column with sidestream drawoff

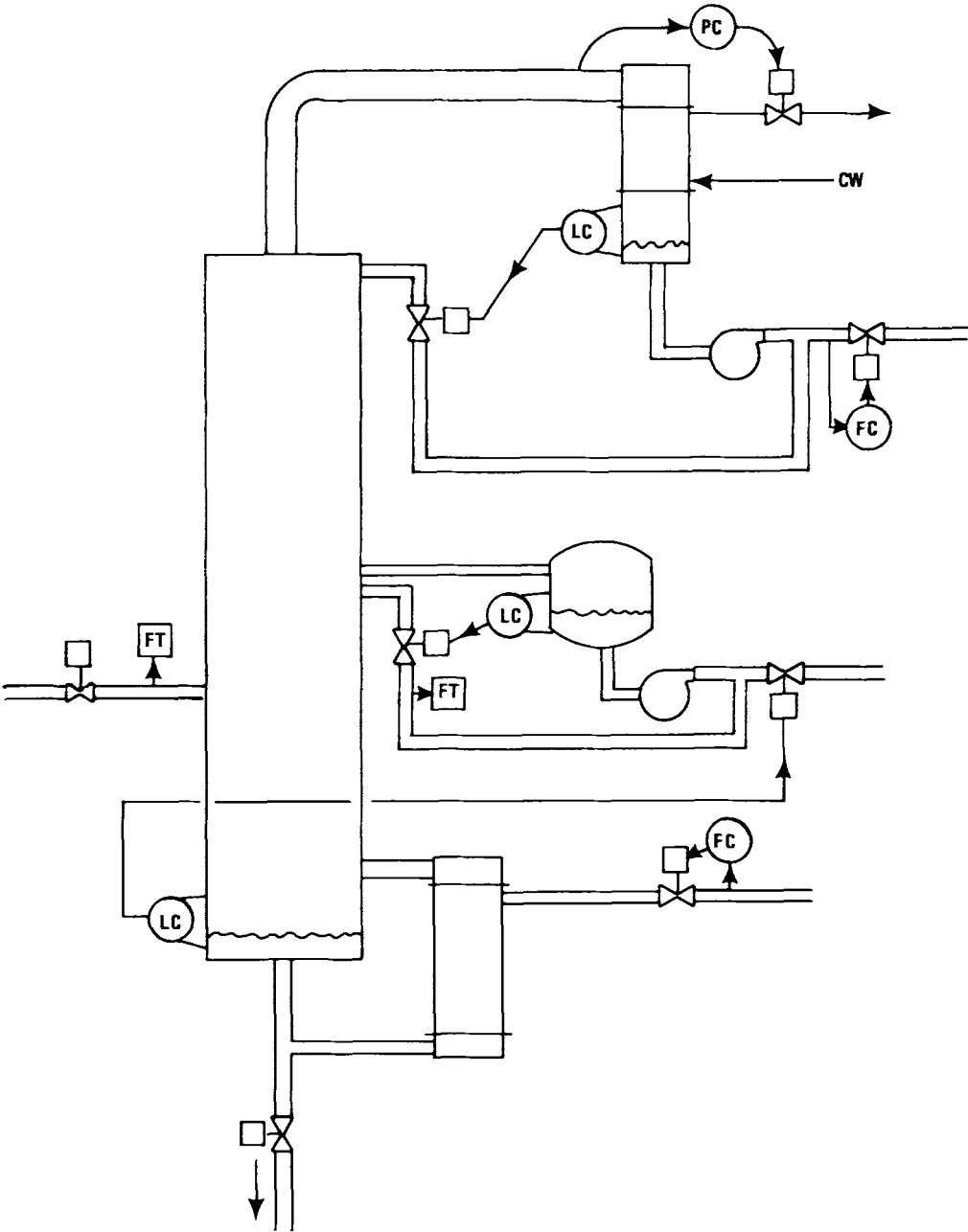


FIGURE 7.2
Controls for liquid sidestream drawoff column

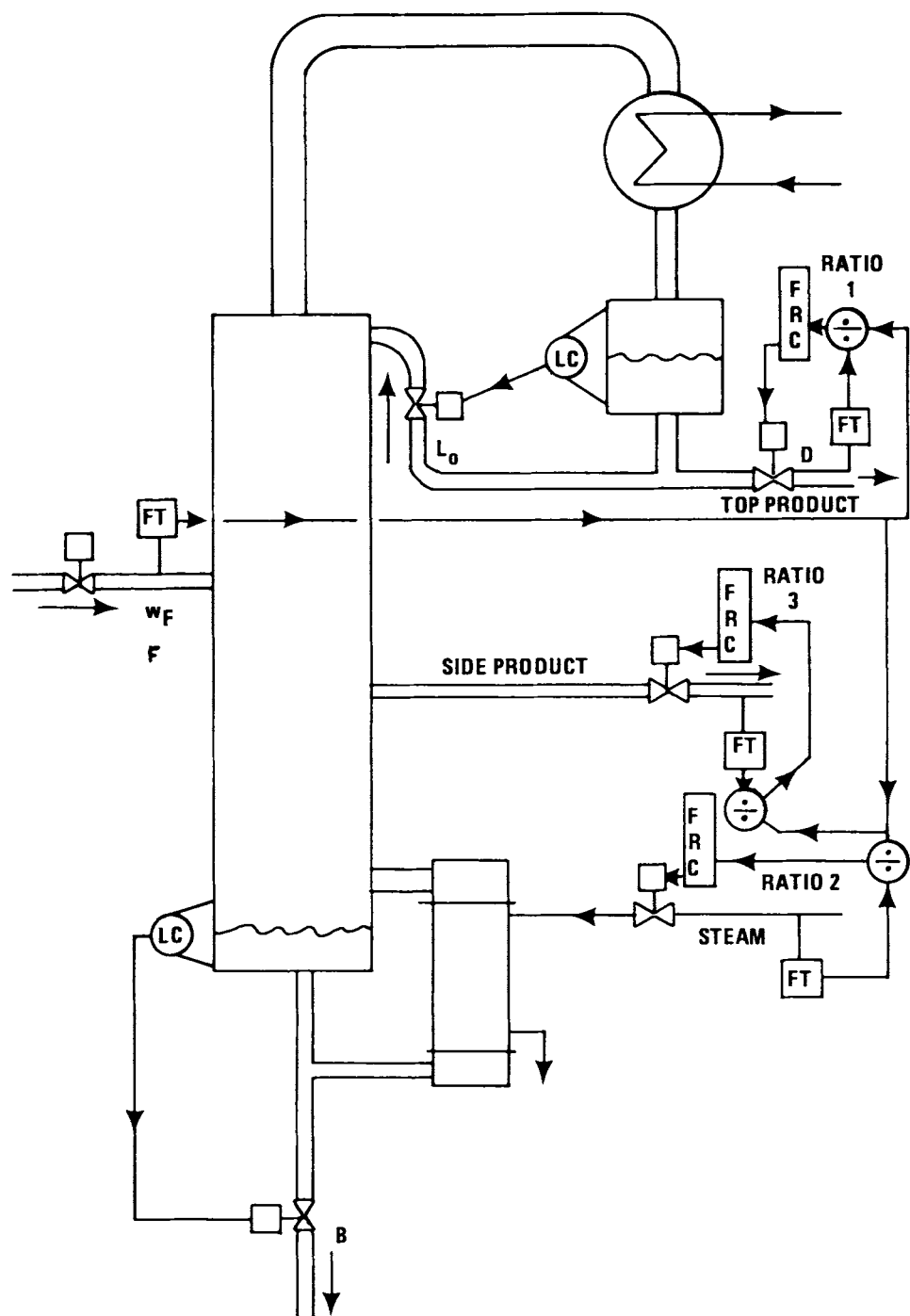


FIGURE 7.3
Alternate control scheme for column with sidestream drawoff

boilers *A*, product *B*, and lumped high boilers *C*. The top-product stream will consist mostly of *A*, the bottom-product stream mostly of *C*, and the side product mostly of *B* with some *A* and *C*.

The following control strategy therefore is suggested:

1. If there is too much *A* in the side product, increase the top-product/feed ratio. See Figure 7.4.
2. If there is too much *C* in the side product, increase the bottom-product/feed ratio. See Figure 7.4.
3. If there is too much side product *B* in either the top or bottom product streams, increase the steam/feed ratio. This increases the reflux ratio, thereby decreasing the concentration of intermediate boiling component *B* at both ends of the column. See Figure 7.5.

Figure 7.5 illustrates a method of minimizing the concentration of *B* in either the top product or bottom product. If the signal from either analyzer becomes too high, steam flow is increased. For pneumatics the two bias relays are each biased such that relay output is 9 psig when component *B* reaches its maximum permissible concentration.

If turndown of more than 2:1 is anticipated, and if composition control is used, impulse feedforward compensation (see Chapter 12) for feed flow will do a better job than ratio controls.

It should be noted that many existing side-draw columns have neither composition controls nor ratio controls. In some cases the purge flows are so small that the loss of product in them is deemed negligible. In view of the rising cost of energy, however, many columns can make good use of steam/feed ratio controls.

7.5 AN IMPROVED APPROACH TO COMPOSITION CONTROL OF SIDE-DRAW COLUMNS

To gain one more degree of freedom for composition control, Doukas and Luyben² came up with the idea of changing the location of the sidestream drawoff tray.

As shown in Figure 7.6A, there are four specifications for the three product streams (three components assumed), which are controlled as follows:

1. The concentration of the intermediate component in the top product (distillate) is controlled by manipulating the reflux ratio.
2. The concentration of the lightest component in the sidestream product is controlled by the location of the sidestream drawoff tray. As shown in Figure 7.6B, this easily can be implemented by conventional analog hardware. The fixed-gain relays are calibrated so that only two of the drawoff valves can be partially opened simultaneously. Thus as the signal from the XS1 composition controller increases, the valves for trays higher in the column open as valves for trays lower in the column close.

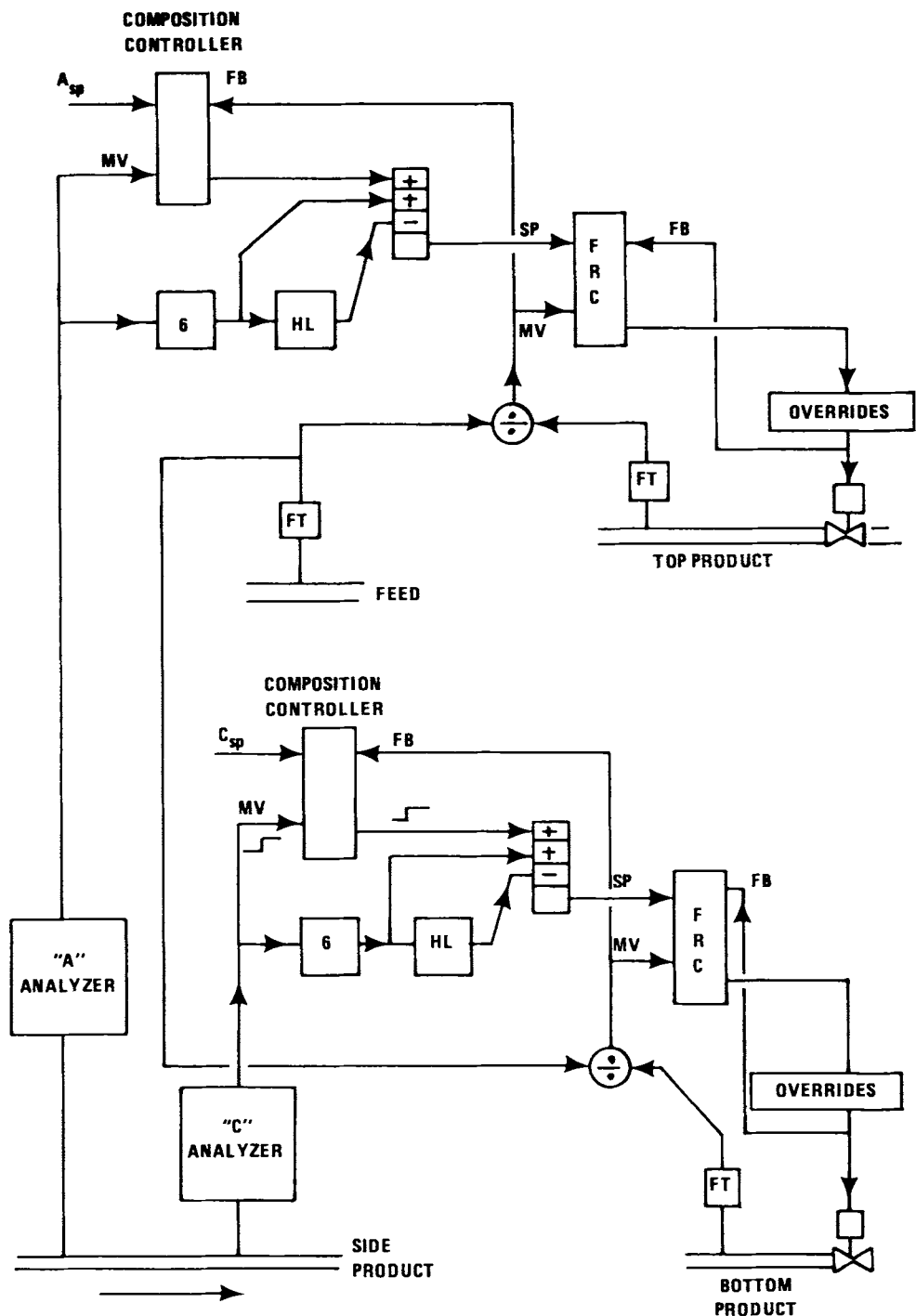
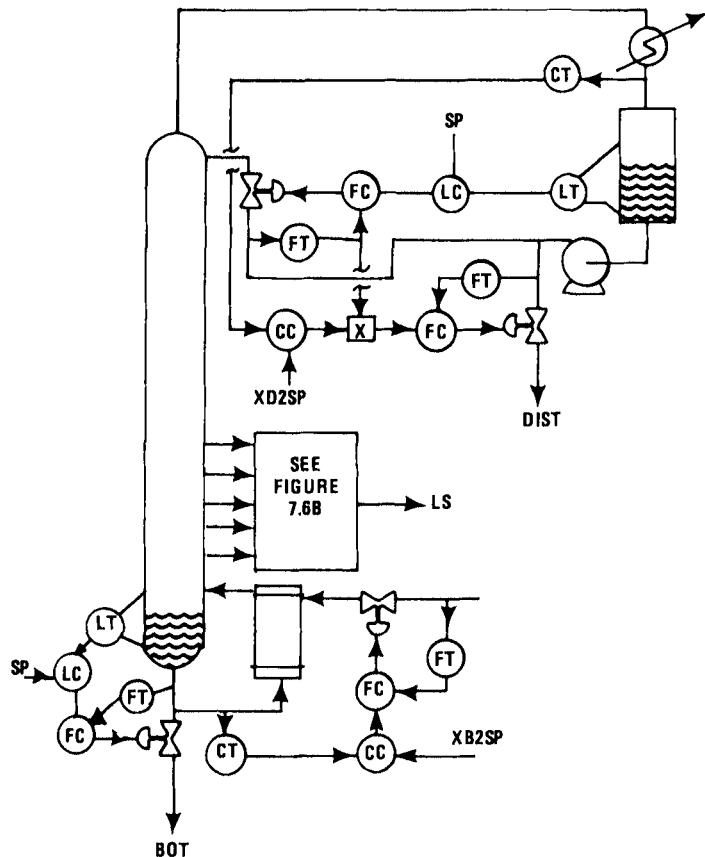
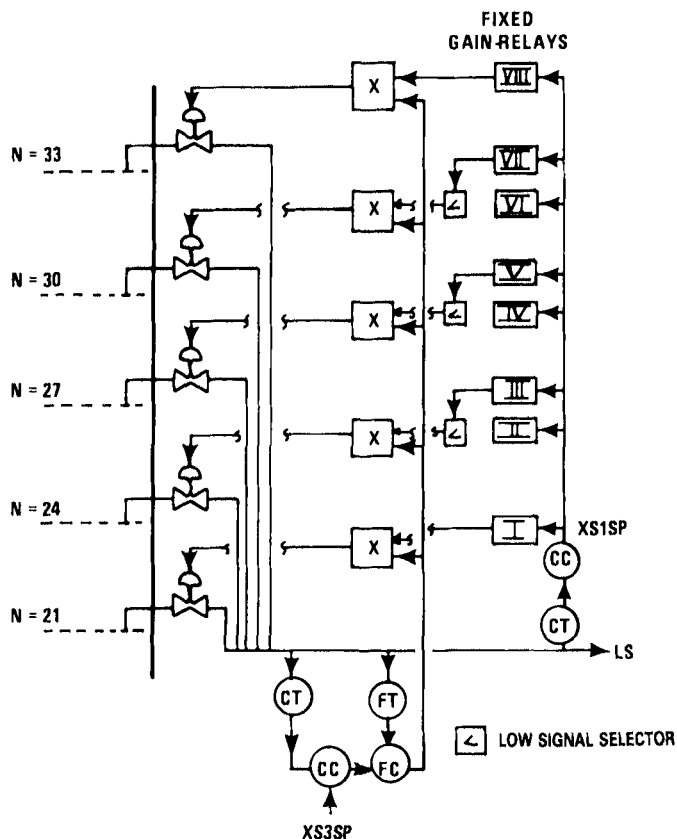


FIGURE 7.4
Scheme for control of sidestream composition



A. OVERHEAD AND BOTTOM CONTROLS



B. SIDESTREAM CONTROLS

FIGURE 7.6

(A) In the control system finally chosen, the toluene impurity content in the distillate product is controlled by the reflux ratio. (B) The five alternative sidestream tray positions and their controls, which regulate the benzene and xylene impurities in the sidestream drawoff, are shown in this blowup.

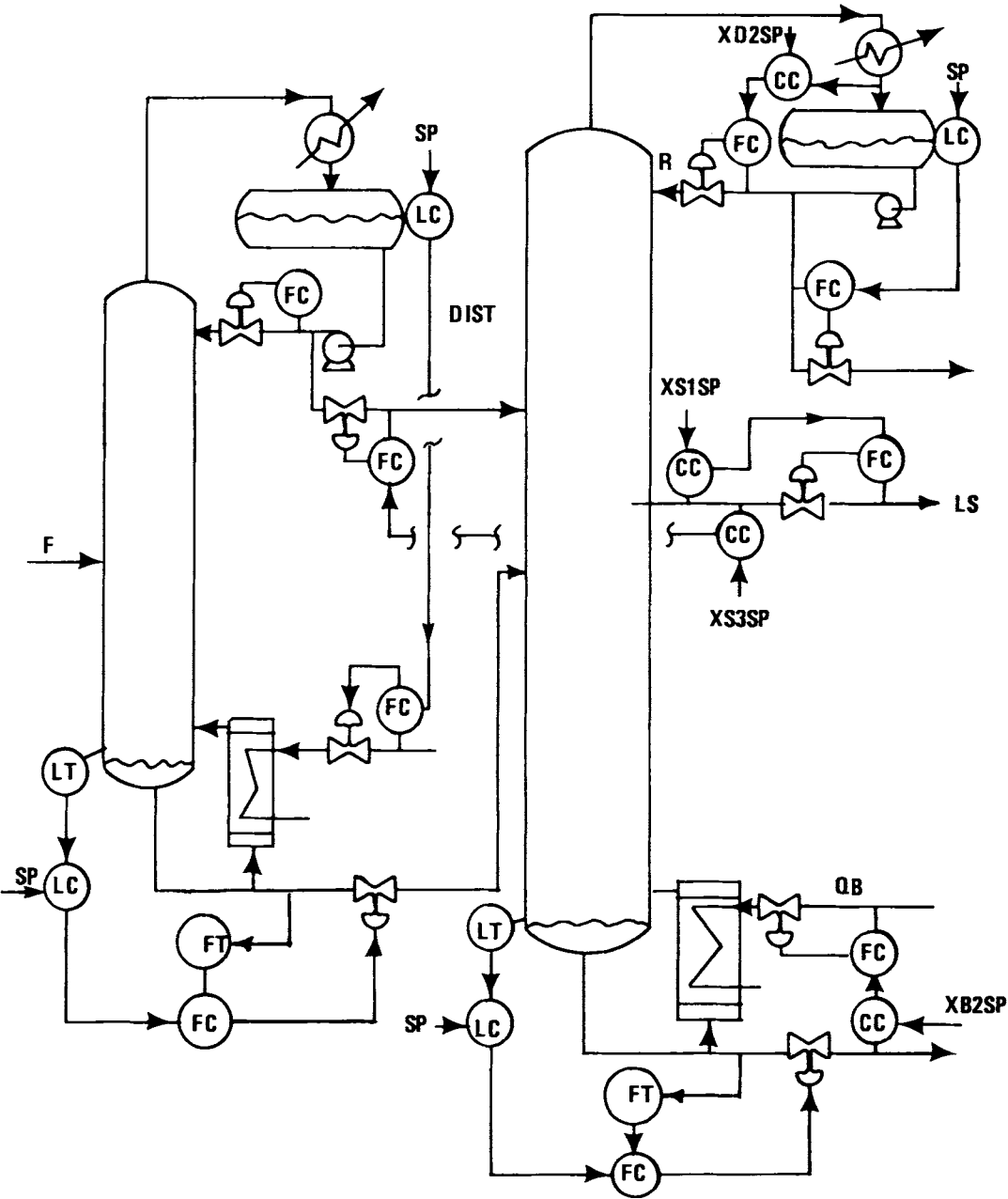


FIGURE 7.7
D-scheme

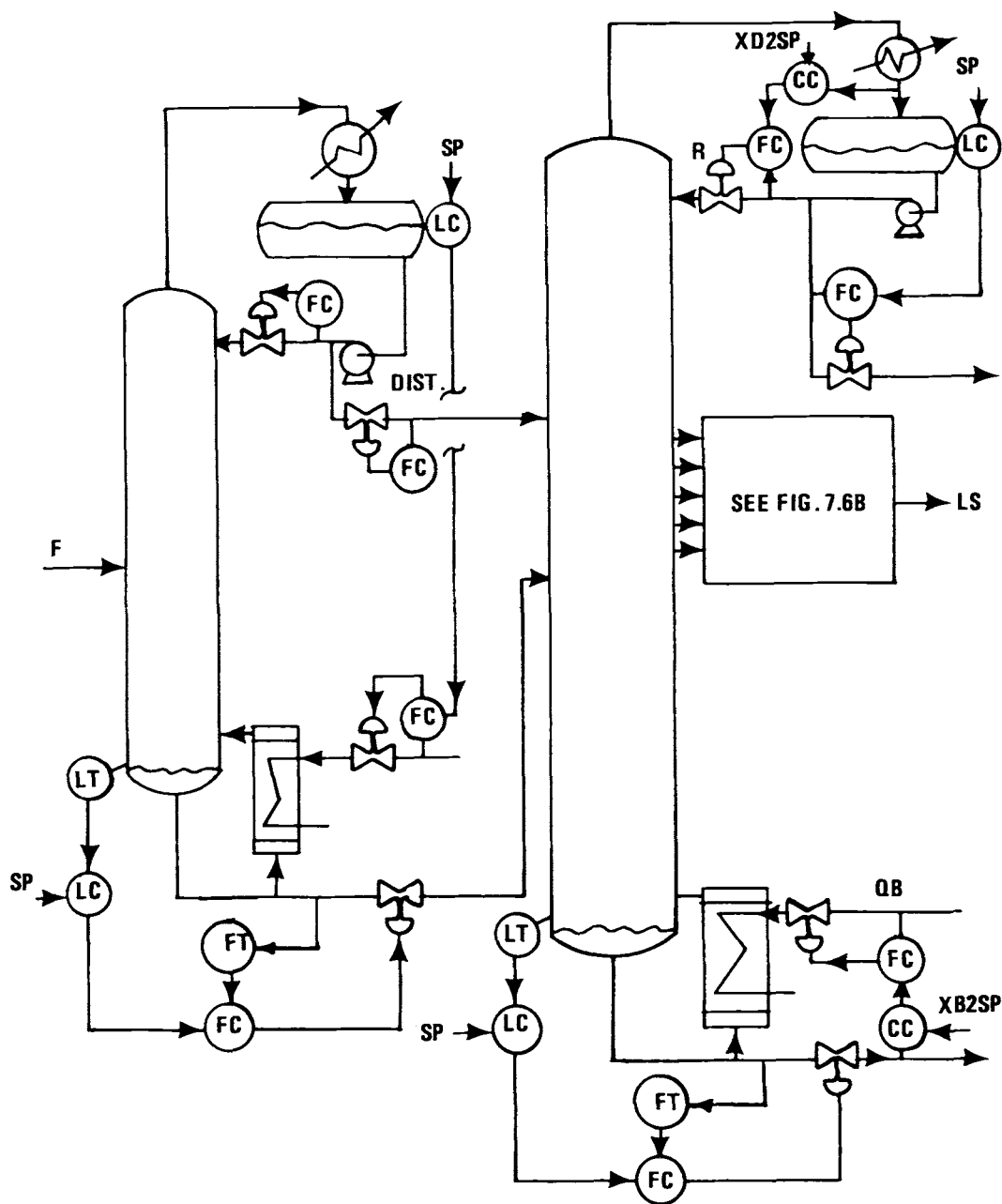


FIGURE 7.8
L-scheme

intermediate boiler, and (2) to remove essentially all of the high boiler out of the bottom, together with the remainder of the intermediate boiler. The two product streams from the prefractionator are fed to two different trays of the sidestream drawoff column. The intermediate product is then withdrawn from a tray or trays located between the two feed trays. Thus the heaviest and lightest components are detoured around the section of the column where the sidestream is withdrawn, and "pinches" are avoided.

Figure 7.7 has been designated by Doukas and Luyben as the "L" scheme. Sidestream drawoff location is used to control the concentration of the lightest component in the sidestream product. The "D" scheme of Figure 7.8 provides manipulation of the distillate product flow from the prefractionator to control sidestream composition.

Both of these schemes were shown to provide effective control for modest changes in feed composition. Since the D scheme is probably easier to implement, it is recommended for most systems.

7.7 OTHER SCHEMES

It is impractical to present here all possible schemes for controlling sidestream drawoff columns. A number of these are discussed briefly in a paper by Luyben.³

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8

Minimizing Energy Requirements

8.1 INTRODUCTION

there are two basic approaches to minimizing distillation column energy requirements:

1. *Conservation*—designing and operating a column so that it makes the specified separation with the least amount of energy per pound of feed.
2. *Energy recovery*—recovering and reusing the heat in the column product streams, whether they be liquid or vapor.

The main emphasis of this chapter is on the latter approach, but it is pointless to try to recover energy unless we also try to conserve it. Consequently we will discuss conservation first.

8.2 CONSERVATION

For distillation, conservation means designing and operating a column so that it makes the specified separation with the least amount of energy per pound of feed. We have a number of techniques to accomplish this:

1. Automatic control of composition of product streams. Operators commonly overreflux conventional columns with a single top product and a single bottom product. Extra heat is used to ensure the meeting or exceeding of specified product purities.

Geyer and Kline¹ give, as an example, a 70-tray column separating a mixture with a relative volatility of 1.4 and with specifications of 98 percent low boilers overhead and 99.6 percent high boilers in the base. If the operator adds enough boilup and reflux to increase overhead purity to 99 percent and base purity to 99.7 percent, an increase of 8 percent in energy consumption results.

2. Feed provided at the proper feed tray. It can be shown that this results in a lower energy requirement per pound of feed than would feeding on any

other tray. As feed composition or enthalpy deviates from design values, the optimum feed-tray location also changes.

3. Column operation at minimum pressure.² Lower pressure usually means higher relative volatility. Therefore, the necessary separations can be accomplished with lower boilup/feed and reflux/feed ratios. Condenser capacity may be limited, however, and the column may flood at lower boilup rates than it would when operating at higher pressures.

4. Use of lowest pressure steam available.¹ In many plants excess low-pressure steam is available that otherwise would be vented to the atmosphere. This steam is usually cheaper than high-pressure steam. Where reboiler ΔT might be too small if the steam were throttled, one may use a partially flooded reboiler (see Chapters 4 and 15) and throttle condensate. Since low-pressure steam is seldom available at constant pressure or steam quality, pressure and temperature compensation of flow measurements is highly desirable if steam is throttled instead of condensate.

5. Use of steam condensate receivers. In many plants steam traps require considerable maintenance and have significant leakage. The use of steam condensate receivers instead of traps reduces maintenance and steam losses.

6. Possible use of mechanical vacuum pumps. For vacuum columns there is some opinion¹ that mechanical vacuum pumps offer energy savings over steam jets. The difference, however, is usually small.

7. Dry distillation. For columns now using live steam, it is sometimes economical to switch to steam-heated reboilers.

8. Insulation. Older columns, designed before the energy crunch, can often benefit from new, increased insulation.

8.3 DESIGN CONSIDERATIONS IN HEAT-RECOVERY SCHEMES

Energy recovery in a distillation column means, practically speaking, recovering or reusing heat contained in the column product streams, whether they are liquid or vapor. A number of schemes have appeared in the literature. The two chief ones involve (1) "multiple effect" distillation, analogous to multiple effect evaporation, and (2) vapor recompression. But, regardless of the scheme, there are five design factors that must be considered:

1. Reserve capacities that may be required:
 - Extra heating capacity
 - Extra cooling capacity
 - Extra distillation capacity

These are important for startups and shutdowns, changes in production rate, changes in feed composition, and changes in product specifications. Generally speaking, however, auxiliary* reboilers and condensers should be avoided, if

* "Auxiliary" condensers and reboilers are those installed in parallel with "normal" condensers and reboilers for startup or peak load purposes.

at all possible. Their use increases investment, as well as instrumentation and control complexity.

Some users have had problems with turning auxiliary condensers and reboilers on and off, and they prefer not to do it. Instead they always maintain at least a small load on these heat exchangers. This obviously wastes energy.

2. Priorities. If recovered energy is to be distributed to several loads, what is the order of priorities?

3. Interactions. Elaborate heat-recovery schemes are often highly interactive; how is this to be dealt with?

4. Overall heat balance. How is this maintained?

5. Inerts (low boiler) balance. With elaborate heat-recovery schemes, this is sometimes a problem. Too high a concentration of inerts or low boilers will blanket process-to-process heat exchangers; too low a concentration will result in product losses through the vents.

In view of the above, it is apparent that control of columns with heat-recovery schemes is more difficult than control of conventional columns.

Let us now look at three types of multiple-effect distillation.

8.4 MULTIPLE LOADS SUPPLIED BY A SINGLE SOURCE

Sometimes, as shown in Figure 8.1, a column that is a very large energy user becomes the energy source for a number of loads, each of which acts as a condenser. Two methods have been used to allocate the energy to be recovered: (1) throttling the vapor-heating medium to each condenser, as shown in Figure 8.1, and (2) operating each condenser partly flooded by throttling the condensate. Some priority scheme must be established for startups and for any other occasion when vapor supply is temporarily short.

One method of handling the priority problems is to use overrides and to split-range the various valves involved, as shown in Figure 8.2. The scheme shown illustrates the use of pneumatic devices, but the concepts readily may be implemented with some digital or analog electronic controls. For the six loads in Figure 8.2, we employ six gain 6 relays. For load 1, which has the highest priority, the gain 6 relay is calibrated to have an output span of 3–15 psig for an input span of 3–5 psig. Load 2 has the next highest priority, so its gain 6 relay is calibrated for 5–7 psig input, 3–15 psig output. This continues until the load with lowest priority, load 6, has a gain 6 relay calibrated for 13–15 psig input, 3–15 psig output. At the design stage of such a system, considerable care must be exerted to obtain suitable value of controller gain and also proper valve sizing.

Process-to-process heat exchangers are commonly designed for very small temperature differences, say 8–10°C. If vapor throttling is used, it should be recognized that the vapor-supply valves will tend to have small pressure drops. Accordingly, it is advisable to have vapor flow control to each load, with a set

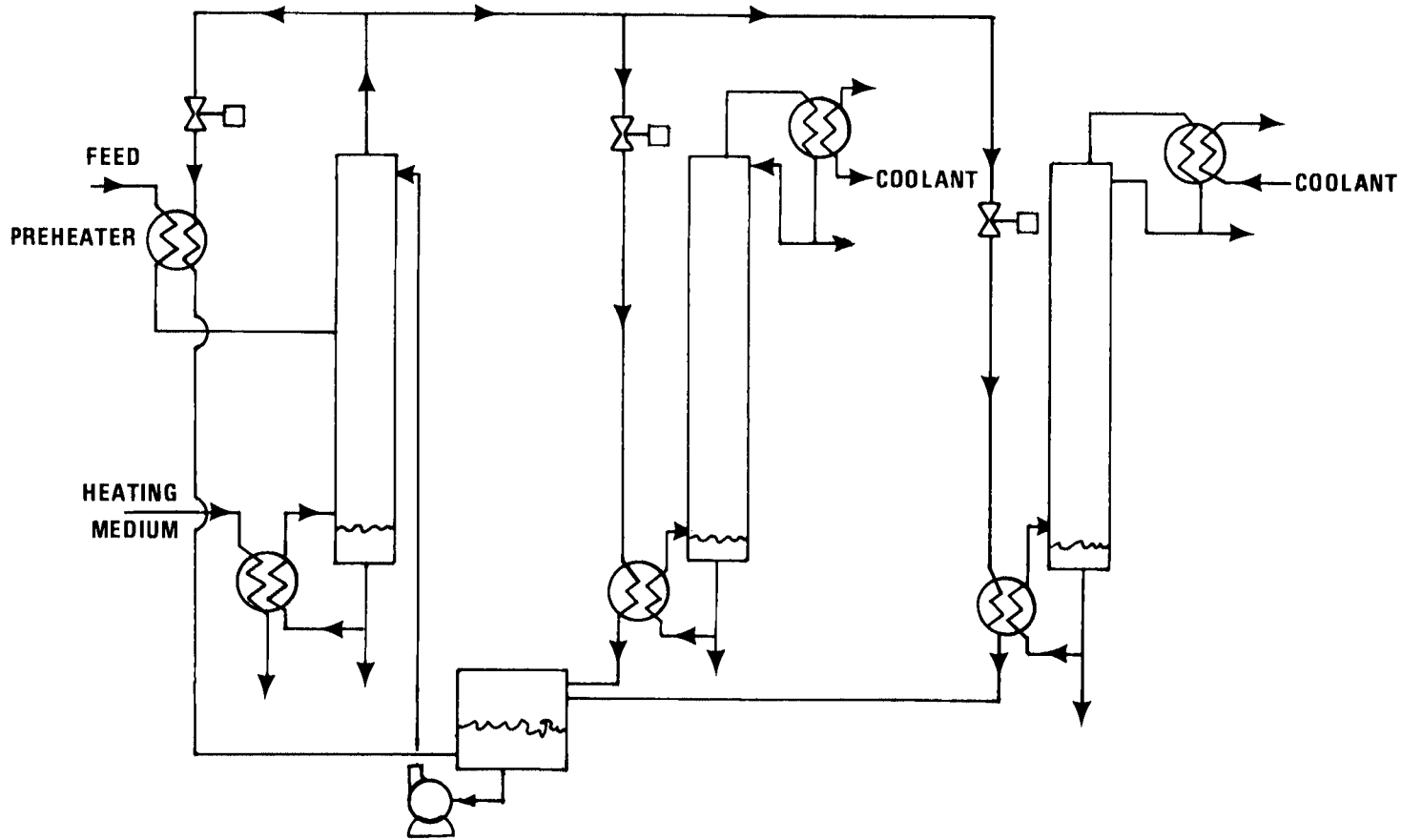


FIGURE 8.1
Heat recovery scheme—single source, multiple loads

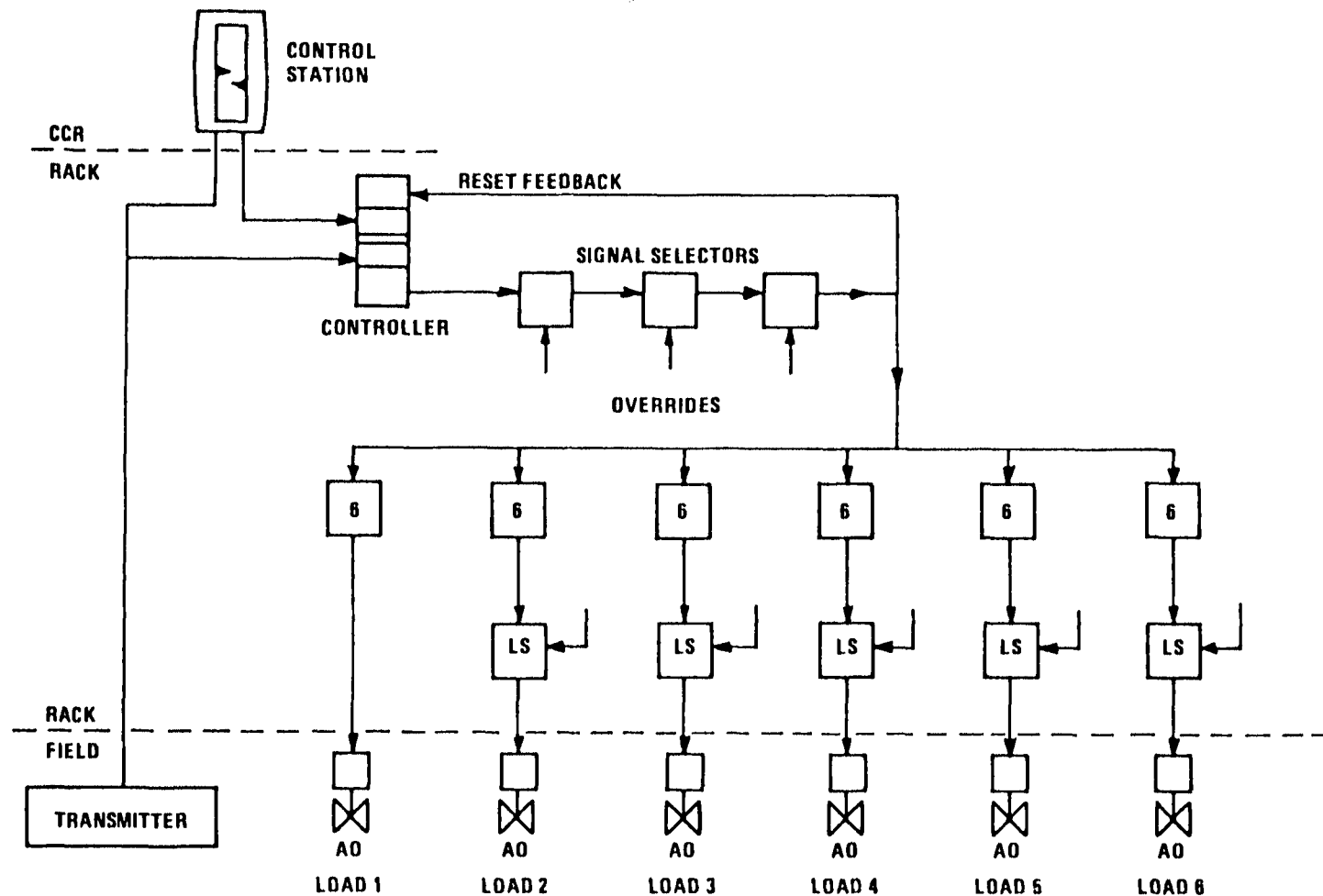


FIGURE 8.2
Scheme for establishing heat load priorities

point from a primary controller. In the absence of vapor flow control, interactions may be severe, and very close control of supply and load pressures may be required.

8.5 SINGLE SOURCE, SINGLE LOAD

When there is only one source and one load (see Figure 8.3), control may be both simpler and more flexible. The column that is the source does not need to be operated at a constant pressure—in the scheme shown, it finds its own pressure. For the illustrative example, the overhead composition of the supply column is controlled via reflux; the base composition of the load column is controlled by boilup in the supply column.

The scheme of Figure 8.3 has an interesting dynamics problem. The controls must be so designed that changes in vapor flow from the supply column must reach the condenser–reboiler at about the same time as feed flow changes from the supply column. If there is a serious discrepancy, particularly if the second-column bottom-product flow is small, base level in the second column may experience serious upsets.

Another problem associated with this scheme is the selection and sizing of the feed valve to the first column. This column will run at a low pressure at low feed rates and at a higher pressure at high feed rates. Assuming that the feed comes from a centrifugal pump, one can see that valve pressure drop will be very high at low flow, and low at high flow. The variation in valve pressure drop with flow will be much greater than that normally encountered in a pumped system.

In another version a following column in the train supplies heat to a preceding column as shown in Figure 8.4. In this particular case, the first column gets only part of its heat from the second column; the remainder comes from an auxiliary reboiler. Interactions between the two columns may be severe. Again, for the cases studied, we have found it advantageous to let pressure find its own level in the second column, that is, the one supplying heat.

An interesting practical problem here is how to adjust the auxiliary reboiler on the first column. After examining some complex heat-balance schemes, we decided that the simplest approach was to use column ΔP . Vapor flow to the first column from the condenser–reboiler will not be constant, but the ΔP control will provide a rapid method of ensuring constant boilup. The ΔP control, in turn, may have its set point adjusted by a composition controller for the lower section of the first column.

It should be noted that for the schemes of both Figure 8.3 and Figure 8.4, maximum column pressure occurs at maximum feed rate and boilup rate. For columns we have studied to date, there has been no problem with flooding at lower rates and pressures.

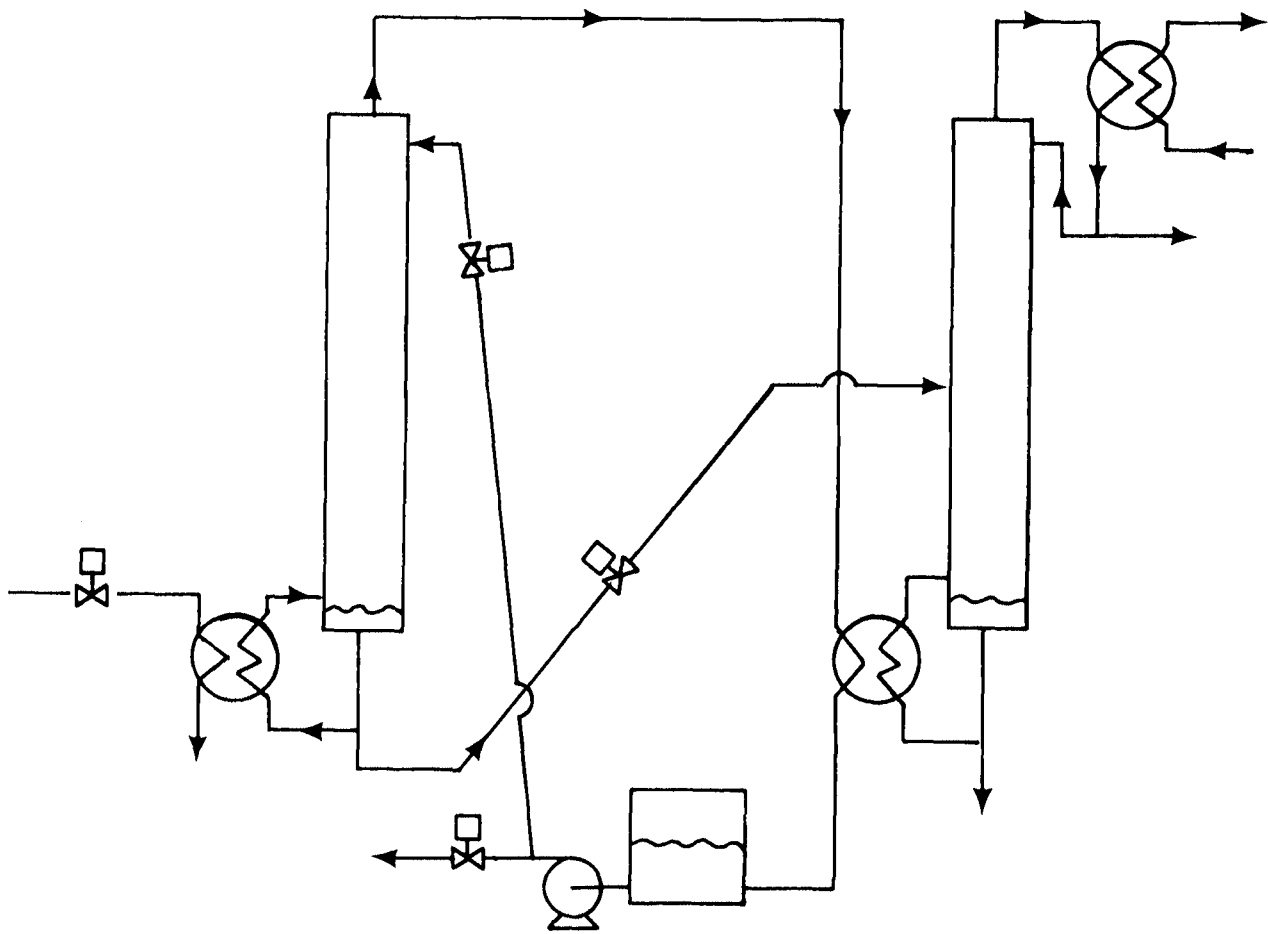


FIGURE 8.3
Heat recovery—single source, single load

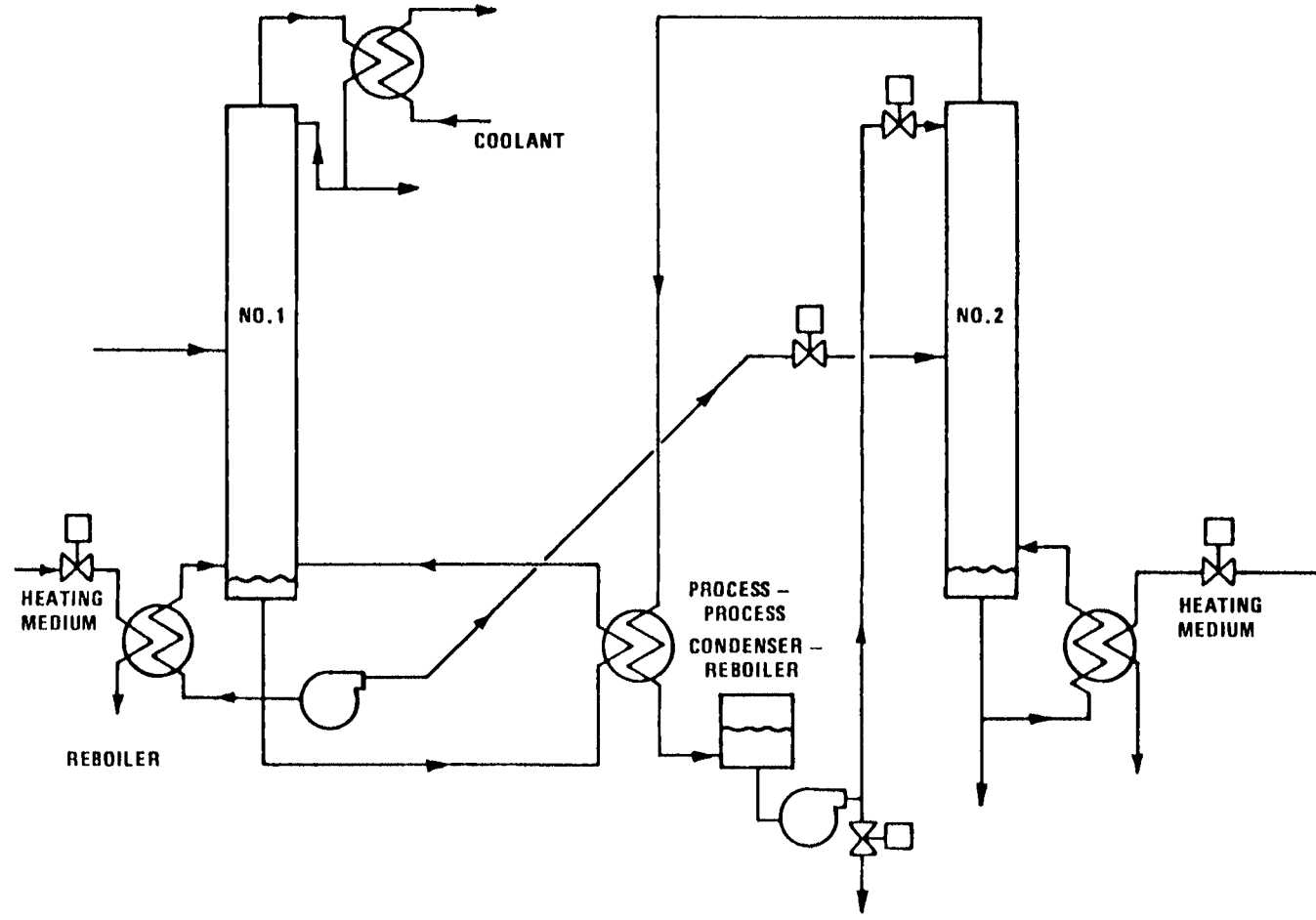


FIGURE 8.4
Heat recovery—single source, single load—scheme 2

8.6 SPLIT FEED COLUMNS

A third arrangement, which is used in some systems,^{3,8} involves splitting feed between two columns that make the same separation (Figure 8.5). The supply column, however, runs at a higher pressure than the load column. The feed split is controlled to maintain a heat balance.

8.7 COMBINED SENSIBLE AND LATENT HEAT RECOVERY

In addition to the recovery of the latent heat of vapor streams, in many cases it is practical to recover part of the sensible heat in the column bottom product and steam condensate by exchange with column feed. Such schemes have been used in the chemical and petroleum industries for years. Since feed flow is typically set by level controllers or flow-ratio controllers, its flow rate will not be constant. The feed enthalpy or temperature, therefore, is apt to be variable. This may make column-composition control difficult unless one employs either feedforward compensation or a trim heater with control for constant temperature or enthalpy. (See Chapters 5 and 11.)

8.8 ENERGY RECOVERY BY VAPOR RECOMPRESSION

In the past vapor recompression ("heat pumps") has often been considered for distillation of materials boiling at low temperatures. The incentive in many instances was to be able to use water-cooled condensers, thus avoiding the expense of refrigeration. Another factor favoring vapor recompression is a small temperature difference between the top and bottom of the column.

Today the main interest is in getting the column vapor compressed to the point where its temperature is high enough to permit using the vapor as a heat source for the reboiler.^{4,5} An auxiliary, steam-heated reboiler and/or auxiliary water-cooled condenser may be necessary for startup (see Figure 8.6). A review of compression equipment and methods of estimating operating costs has been presented by Beesley and Rhinesmith.⁶ Mosler⁷ discusses the control of a number of vapor recompression schemes. Null⁹ presents investment equations and data. Fahmi and Mostafa¹⁰ indicate that the optimum location at which to use the compressed vapor may not be in a reboiler at the column base, but rather at an intermediate site.

Other papers on energy integration for distillation columns include those by O'Brien¹¹ and by Rathore et al.¹²

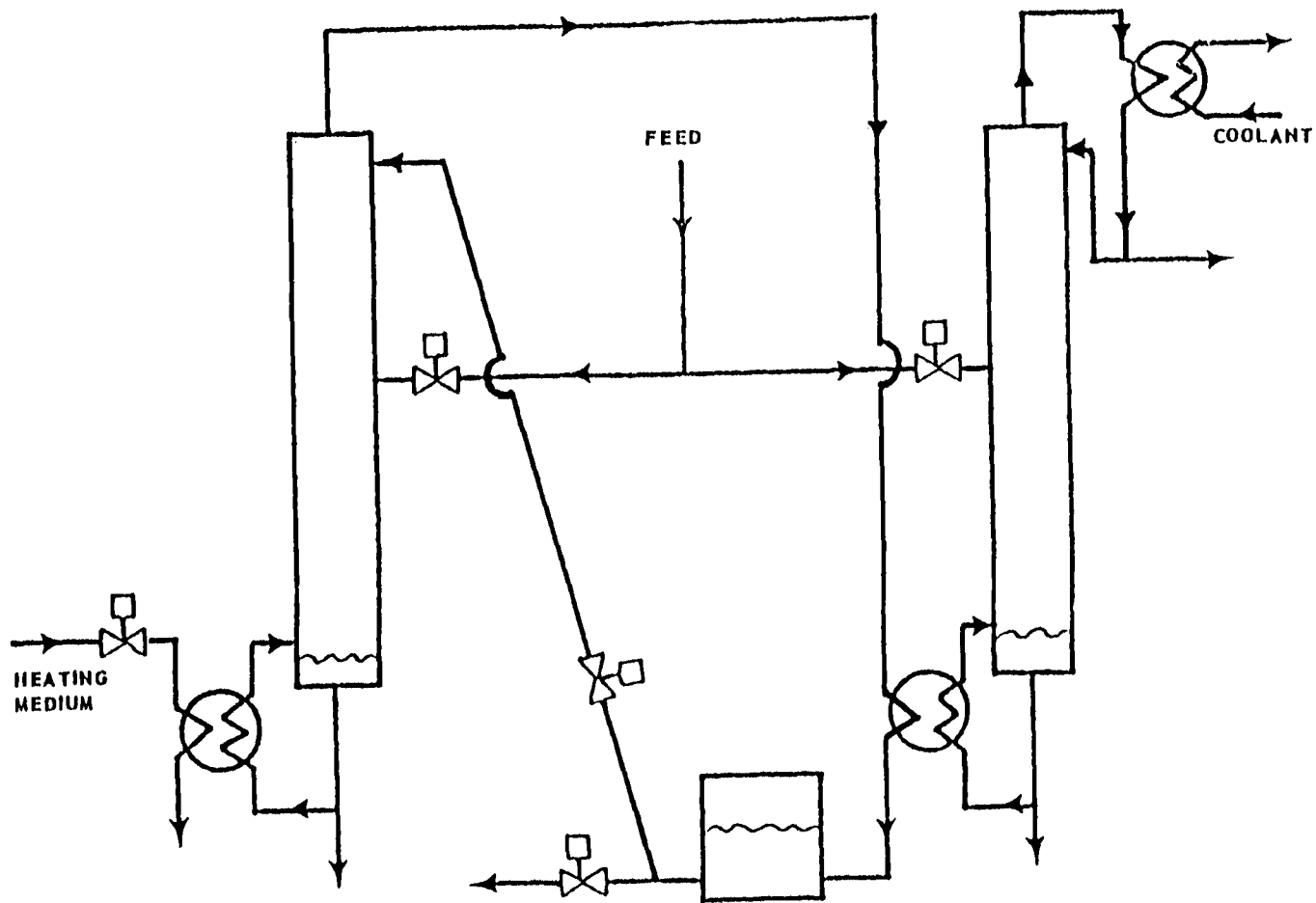


FIGURE 8.5
Heat recovery—split feed, single load

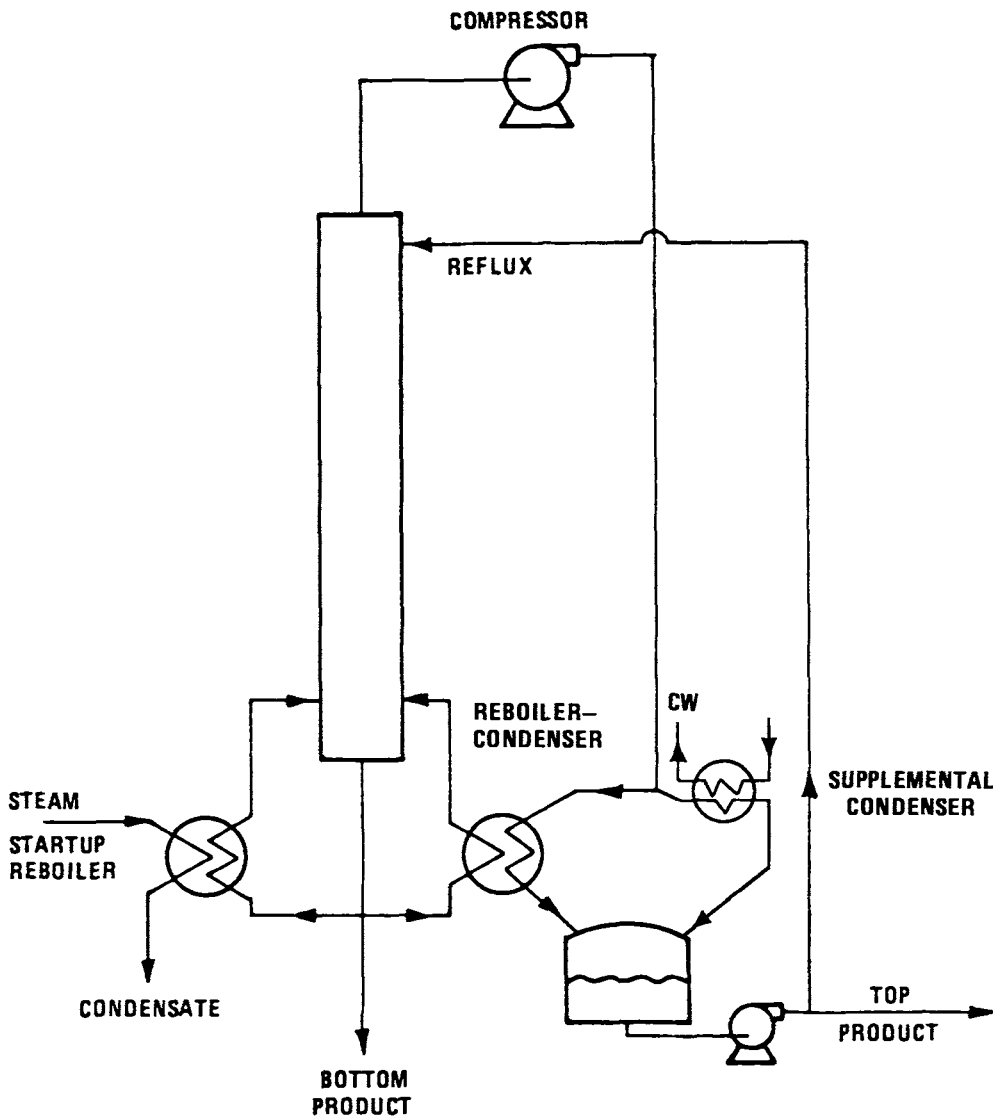


FIGURE 8.6
Heat recovery via vapor recompression

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9 Application of Protective Controls to Distillation Columns

9.1 INTRODUCTION

as mentioned earlier, most existing plant instrument systems consist of a large number of single-loop controls: one transmitter, one controller (usually proportional-reset), one manual/automatic station, and one valve. In real life, however, many a valve must be adjusted in accordance with changes in more than one process variable. For example, the steam valve to a distillation column reboiler may have to be adjusted in response to as many as half a dozen variables. With existing systems the operator usually must put the control loop into the “manual” mode and adjust the control valve position as all but one of the pertinent variables change. To put it another way, conventional “automatic” controls are automatic only in a limited sense over a limited range of conditions. One must superimpose upon them a kind of logic that spells out various courses of action to take as certain constraints, such as high column ΔP , low tank level, or high temperature, are approached or reached. For conventionally instrumented plants, this logic is a major part of the formal instructions for operating personnel. In the case of highly integrated (“bootstrap”) plants, the required logic, although it may be deduced and written down in advance, may be too detailed and complex to be absorbed readily by human operators.

In addition, the required reaction times in certain circumstances may be too short for typical human physiology. Automatic controls, on the other hand, provide continuous (or almost so for digital controls), simultaneous surveillance and recognition of operating conditions and continuous, quantitative response to them. They are also much more reliable than human beings.

To make a given control valve respond to more than one controller, we must have a means of telling the valve which controller to obey. To put it in

process control language, we want a multivariable control system instead of multiple, single-variable control loops. In the remainder of this section, we illustrate a number of techniques for accomplishing this with specific applications to distillation columns. The usual objective is to provide protective controls that permit the column to operate close to constraints without exceeding them. In so doing, we greatly improve our ability to achieve certain other control objectives:

- Automatic, or at least easy, startup and shutdown
- Automatic total reflux operation under specified circumstances
- Maximum-capacity operation
- Ability to make desired transitions in column terminal product compositions with minimum production of off-specification products
- Minimum interlock shutdowns, that is, ability to stay on the line a higher percentage of the time
- Minimum turndown requirements for both process streams and utilities

The fourth item is particularly important when feed-stock composition varies widely and it is desired to optimize column or train operation as, for example, with a computer.

We use the term *override* for the most part to refer to the use of two or more controllers connected to a control valve through high or low signal selectors. Logic is built in to enable one controller to “override” (i.e., take over from) the other controller or controllers^{2,3}

9.2 OVERRIDES AND INTERLOCKS

In the chemical and petroleum industries, the most common types of protective controls are interlocks and overrides. Interlocks normally function in an abrupt manner to shut down a piece of equipment or one or more steps in a process. For example, if a column feed pump fails, a low feed flow interlock may be used to shut the column drawoff valves, and perhaps also to shut off steam. High column base temperature and high column differential pressure are sometimes interlocked to shut off steam to the reboiler. Usually interlocks must be reset manually by the operator.

By contrast, overrides can be designed to provide gradual, rather than abrupt, corrective action and function in both directions, that is, they do not have to be reset manually. As a generalization, interlocks are most useful in cases of equipment malfunction or failure; overrides are most useful in protecting the process and keeping it running as certain maximum or minimum permissible operating conditions are approached.

9.3 IMPLEMENTATION OF OVERRIDES

Many ways of implementing overrides are possible, but a particular one that is inexpensive and works well involves the use of combinations of devices such as described in the following.

High Selectors (HS) or Low Selectors (LS)

These devices select either the higher or the lower of two input signals. As many input signals as desired may be accommodated by arranging selectors in series. Multiple-input low-signal selectors are now available from vendors. In practice one or more selectors are inserted between the output of a “normal” controller and its final control element, usually a valve. The outputs of the “override” controllers are also connected to these selectors. As process constraints are approached, one of the override controllers will “take over” or “override” the normal controller and drive the final control element in the proper direction—either to force the process away from a constraint, or to hold it a safe distance from a constraint.

One may also construct a median selector with two high selectors and two low selectors as shown in Figure 9.1. The major application has been for auditing multiple flow measurements to a chemical reactor. If either the high or low measurement deviates too far from the median value, an alarm or interlock is activated. So far we have found no applications to distillation, but it is a technique that is worth keeping in mind.

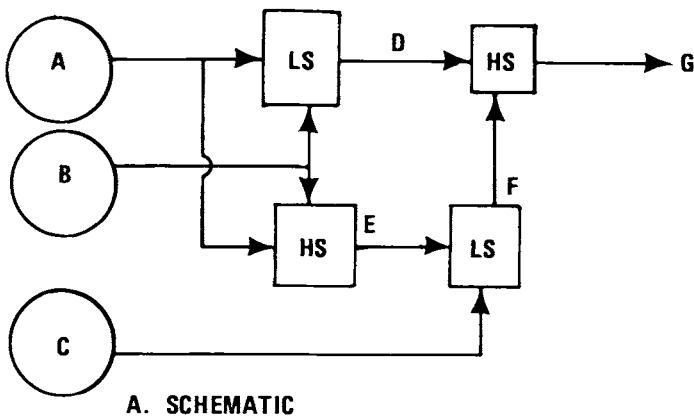
High Limiters (HL) or Low Limiters (LL)

High and low limiters are devices that reproduce the input signal 1:1 up to a predetermined maximum value (*HL*) as shown by Figure 9.2, or down to a predetermined minimum value (*LL*) as shown by Figure 9.3. In either case the cutoff or limiting value is readily adjustable.

Functionally, a high limiter is a low selector plus a signal source such as a supply regulator (Figure 9.4). It thus can be constructed with two separate devices or assembled into one housing. Correspondingly, a low limiter is a signal source plus a high selector (Figure 9.5); it also may be assembled with two devices or purchased as a combined device.

Summers

Summers are devices that add or subtract. If pneumatic versions with adjustable gain are required, one should avoid those with pressure-dividing networks. Their accuracy is poor.



A	B	C	D	E	F	G
1	2	3	1	2	2	2
1	3	2	1	3	2	2
2	1	3	1	2	2	2
2	3	1	2	3	1	2
3	2	1	2	3	1	2
3	1	2	1	3	2	2

B. LOGIC TABLE

FIGURE 9.1
Median selector (J. P. Shunta design)

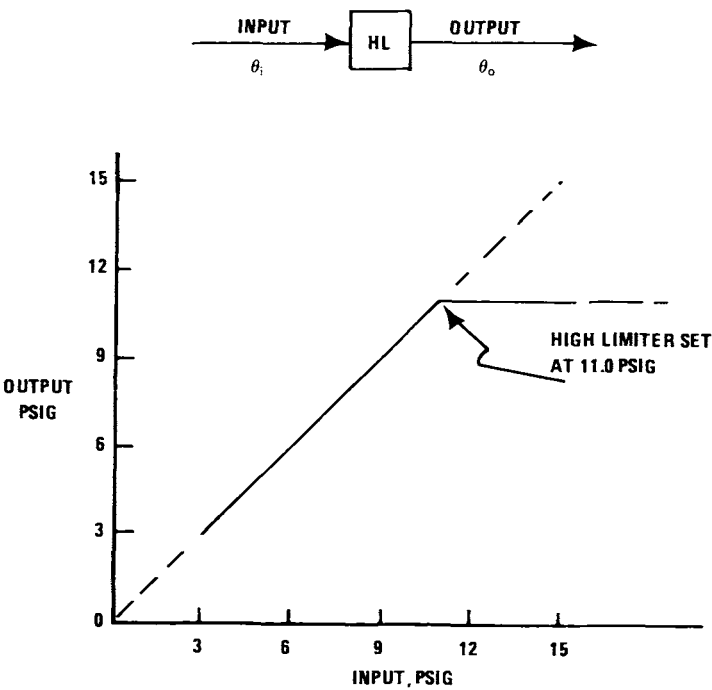


FIGURE 9.2
High limiter

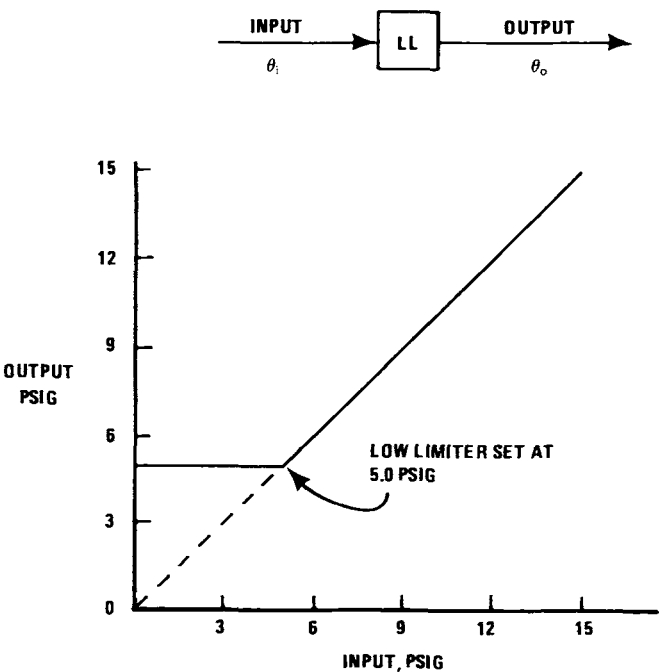


FIGURE 9.3
Low limiter

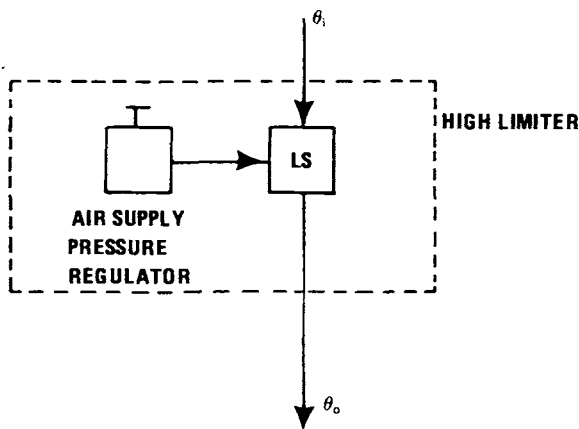


FIGURE 9.4
High limiter schematic

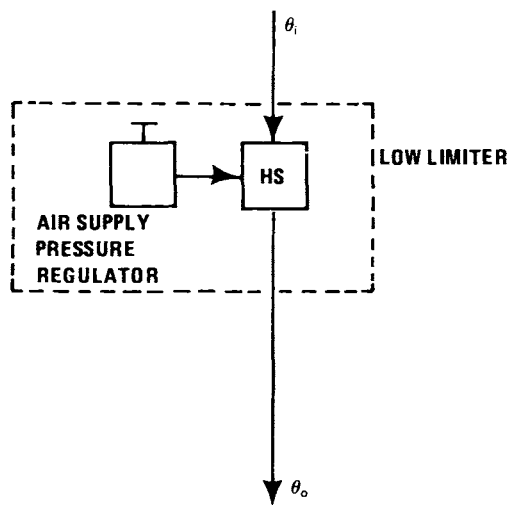


FIGURE 9.5
Low limiter schematic

9.4 CONTROLLERS IN OVERRIDE CIRCUITS

Three types of controllers have been used in override circuits; two of them widely, and the third less frequently.

Proportional-Reset (PI) or Proportional-Reset-Derivative (PID) Controllers

Both of these are commonly used as normal controllers, although the PI type is occasionally used as an override controller. PID controllers, if used, should be so designed that the derivative acts only on the measurement signal, not on the controller output. Both PI and PID controllers must contend with the problem of reset windup, discussed below.

Proportional-Only Controllers

The fixed-gain proportional-only relay, either direct acting or reverse acting, with adjustable bias has become one of the most widely used devices in override circuits. Common values of gain are 2, 3, 4, 6, and 25. Use of the minus sign (–) with these figures implies reverse action, or negative gains.

A proportional-only controller or relay follows a straight-line equation.

$$\theta_o = K\theta_i + B \quad (9.1)$$

where

θ_o = output signal

θ_i = input signal

B = bias

K = proportional gain

We have found it convenient to calibrate pneumatic, fixed-gain relays in terms of the input signal, $[\theta_i]_9$, required to produce 9.0-psig output. Then

$$9.0 = K[\theta_i]_9 + B \quad (9.2)$$

or

$$B = 9.0 - K[\theta_i]_9 \quad (9.3)$$

On substituting equation (9.3) into equation (9.1) we obtain:

$$\theta_o = K(\theta_i - [\theta_i]_9) + 9 \quad (9.4)$$

The instrument may now be calibrated by putting $[\theta_i]_9$ into the input and adjusting the bias for 9.0-psig output.

For override purposes we use proportional-only controllers if possible, largely to avoid some problems associated with PI controllers. If the latter are used for override purposes, no fixed relationship will exist between controller output

and process variable. A sudden disturbance can cause an overshoot above the set point, with the amount depending on the magnitude and rapidity of the upset and the reset time and proportional band of the controller. Further, for a given upset, the output of a proportional-reset controller usually swings through wider limits than the output of a proportional-only controller. These features make this type of override highly undesirable in those applications where, for safety's sake, certain limits must not be exceeded. Maximum temperature in some chemical reactors is a good example of this. We thus use proportional-reset override controllers only in those cases where good control could not be obtained with gain 2 or higher proportional-only overrides because of closed-loop stability problems. Examples include column ΔP or base pressure.

For predetermined maximum and minimum limits, we select a proportional-only controller with a gain that will drive the valve from 3 to 15 psig (or vice versa). Thus:

$$\text{Proportional gain} = \frac{15 - 3}{\text{max} - \text{min}}$$

where "max" and "min" are in terms of the process transmitter output.

Consider a system where a low base level override is to be used to close the steam valve. Its output is compared with that of the normal controller through a low selector. Let us say that we want full override action to occur between zero level (3 psig) and the 25% level (6 psig). Then the required proportional gain is:

$$\text{Proportional gain} = \frac{15 - 3}{6 - 3} = \frac{12}{3} = 4$$

Since the gain 4 relay is to put out 15 psig at the 25 percent level, and its output goes to a low signal selector, it clearly will exercise no control action at a level of 25 percent. On the other hand, if level is dropping, the output of the gain 4 relay will drop below that of other controllers at some value of level above zero. We are never sure, therefore, at what point the low-level override will take over, but we know positively that it will be between the zero and 25 percent levels.

Floating or Integral Controller

This type sometimes is used as a normal controller instead of a PI controller when required proportional gain would be very low or where it would need to be changed frequently as process conditions change.

9.5 ANTI RESET-WINDUP

Single-Loop Controls

When control of a valve is transferred from one controller to another through a selector, we would like this to happen without a bump. Since the

selectors will switch on very small differential signals, they will in themselves introduce no significant jolt. The major potential source of bump is reset windup in overridden controllers.

In all commercial pneumatic controllers, reset action is obtained by a positive-feedback, unity-gain circuit that feeds the controller output signal back through a needle valve into a reset chamber. This feedback is normally internal, but in some commercial controllers it may be taken from an external connection by proper orientation of a switchplate on the body of the controller. If, now, we use the valve-loading signal as the reset-feedback signal, we will get normal reset action when the controller is controlling. When, however, another controller takes over, reset action in the first controller ceases, but its output goes up and down with the valve-loading signal. This tracking action is delayed by the reset time constant, which for many controls (such as liquid flow control) is small. The output of the overridden controller will differ from the external reset feedback signal by the product of its gain, K_c , and the error signal, ε .

The anti reset-windup technique discussed above is known as “external reset feedback.” For most applications either it, or the modification mentioned below, is our preferred scheme. It has the disadvantage that the controller output signal, commonly labeled “valve position,” is really different from the actual position. It differs by the product of the error signal times the proportional gain. Lag in the reset circuit may cause further error. A modification therefore is introduced by some vendors, particularly in the newer microprocessor controls. This consists of setting the reset time equal to zero when the controller is overridden.⁵ This technique is sometimes called “integral tracking.” It should *not* be used with auto overrides.

Another technique is called “output tracking”; the overridden controller output is driven to almost the same value as that of the overriding signal. This does have the advantage that the controller output signal is nearly equal to the valve-loading signal. However, according to Giles and Gaines,⁵ this technique does not work as well, at least under some circumstances, as does integral tracking.⁵ Various other techniques have been described by Khanderia and Luyben.⁶ As an example consider the simple system of Figure 9.6. It shows a distillation column with a base temperature controller and a column ΔP controller, both connected through a low selector to the steam valve, which has air-to-open (AO) action. Let us assume that at startup time the base of the column has enough low boilers that it will boil at a temperature lower than the normal temperature controller set point. If we do not lower the set point, the temperature controller will open the steam valve wide, the column-base contents will boil very rapidly, and column pressure drop will shoot up. As it goes above the ΔP controller set point, this controller output starts to decrease, and when it becomes lower than the output of the temperature controller, it takes over the steam valve through the low selector. The steam valve is now held open just far enough to keep column ΔP at the set-point value (chosen to be the maximum acceptable). Eventually the low boilers are taken overhead and base temperature rises to the point where the temperature controller takes over. Since the operator

does not need to change the ΔP controller set point, it can be located away from the panel in an override cabinet.

Several points should be noted about this system:

—Since the override controller is of the proportional-reset type, there will not exist a fixed, known relationship between the controller output and process transmitter output.

—The overriding controller—in this case differential pressure—must have a smaller reset time than the normal controller or it will sometimes take over at values of column ΔP different from the override controller set point.

—Manual-automatic switching is not necessary since both controllers have anti reset-windup.

—If the system of Figure 9.6 is started up in the automatic mode, the control valve will open and the controlled variable will rise toward the set point at a rate determined by the reset time constant.

The first two points lead us to use proportional-only overrides where possible, while the third suggests that automatic startup and shutdown can be obtained with an air switch and several three-way pneumatic valves.

It should be noted that only a few of the electronic analog controllers on the market have external reset feedback or some other satisfactory anti reset-windup scheme. But newer digital and microprocessor-based controls (as of late 1983) use a variety of techniques, most of which appear to be satisfactory.

Cascade Controls

The preceding technique works well for conventional single-loop controls and for secondary or slave loops in a cascade system. But for primary or master controllers, we do something different since the valve-loading signal is no longer meaningful for reset feedback.

To eliminate reset windup, we break the master controller internal feedback as before, but now we use the secondary measurement for feedback as shown by Figure 9.7. If, for example, we have temperature cascaded to flow, we feed the output from the flow transmitter back into the master controller reset circuit. This means that during normal control the lags in the secondary control loop appear in the reset feedback circuit of the primary controller. If, as usual, the slave loop is much faster than the master loop, this technique will not appreciably increase the master controller reset time.

9.6 FEEDFORWARD COMPENSATION WITH OVERRIDES

A convenient method for providing feedforward compensation that does not interfere with either normal reset or antireset windup involves the use of an “impulse” relay and a summer as shown in Figure 9.8. In pneumatics these functions are sometimes combined into a single device. The impulse function

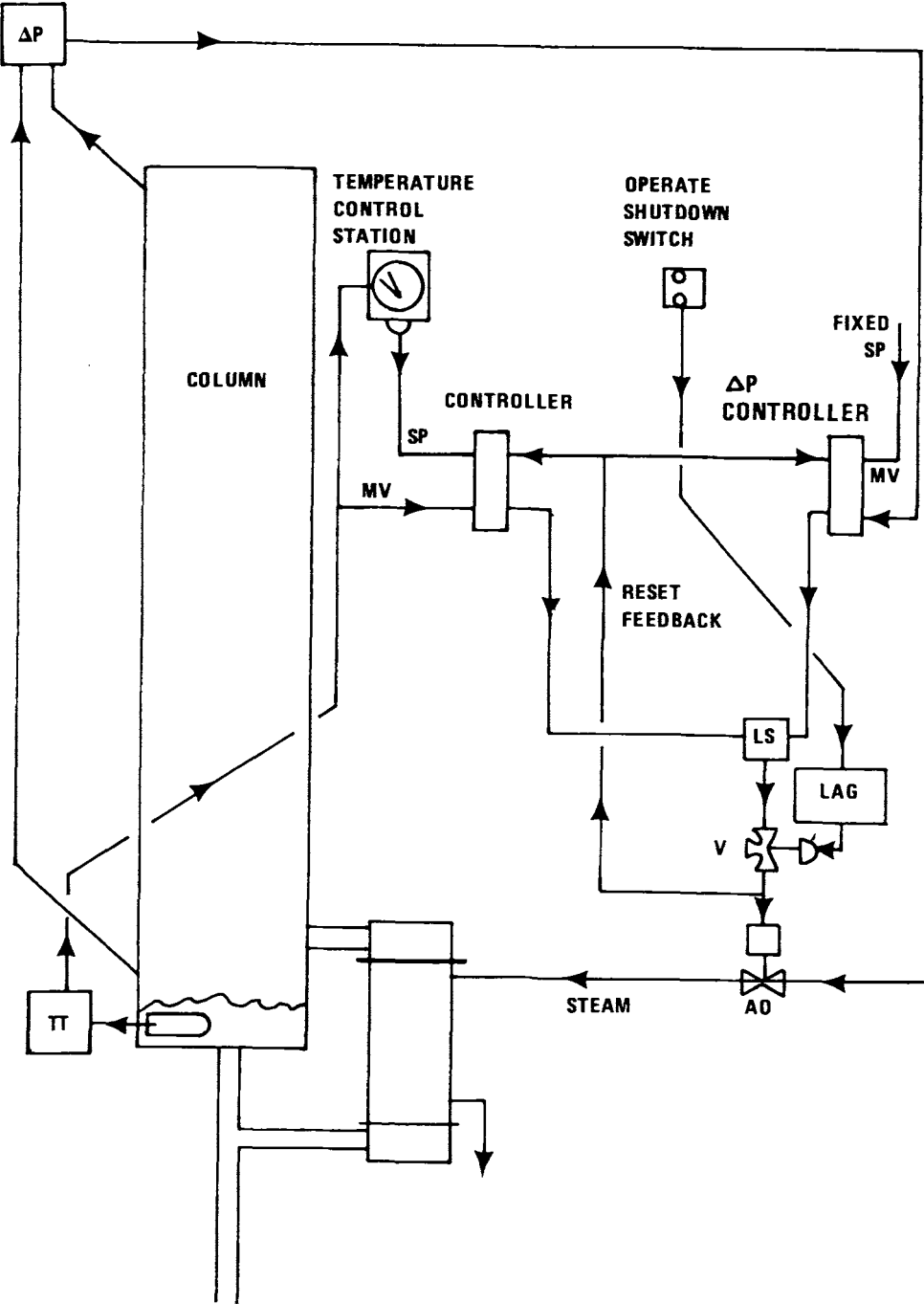


FIGURE 9.6
Column base temperature control with ΔP override

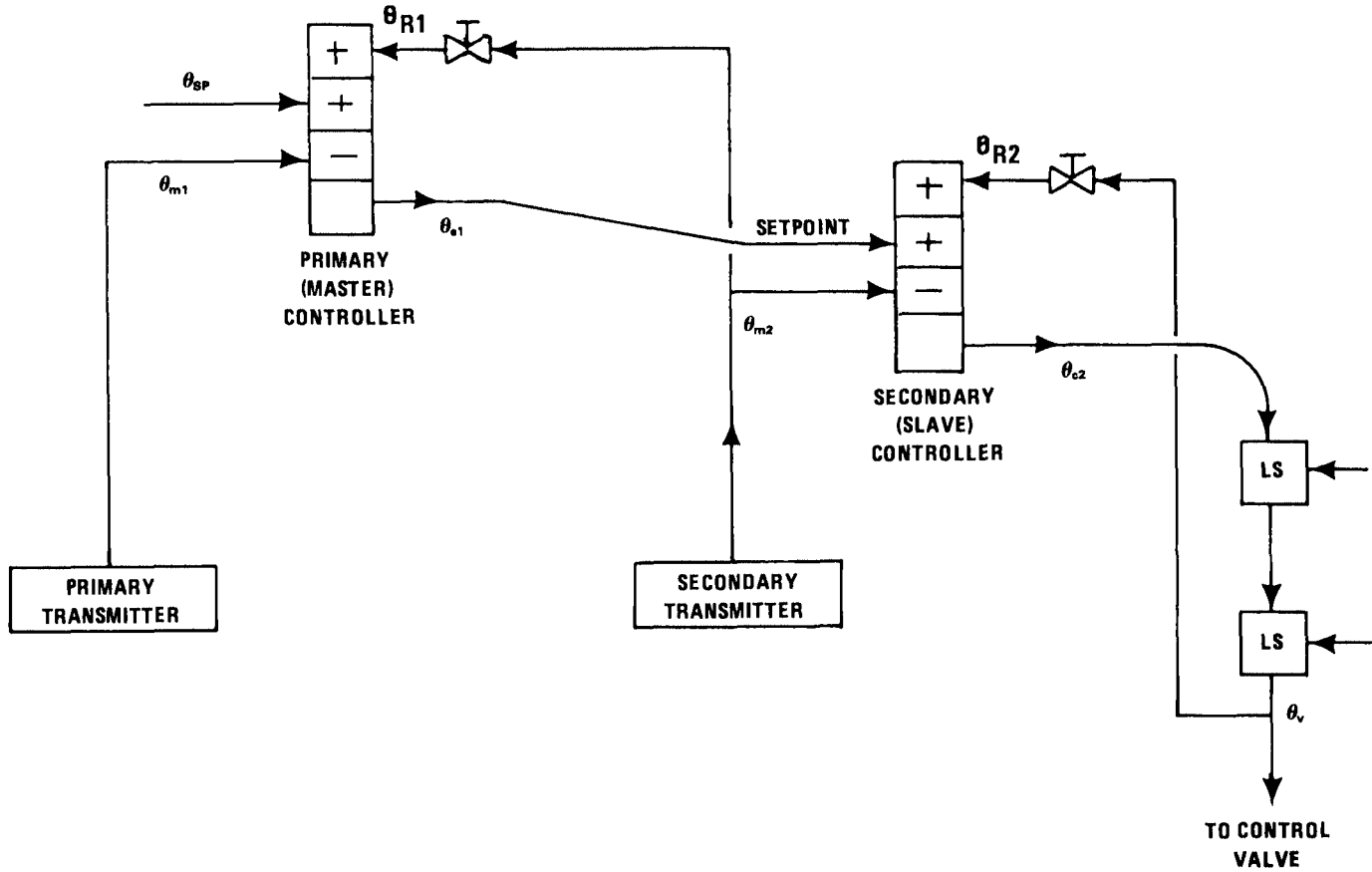


FIGURE 9.7
Antireset windup for cascade loops

passes only the transient part of the feedforward signal. This eliminates or avoids scaling problems with the summer since the steady-state or dc component of the feedforward signal is blocked. It can be shown that best results are usually achieved if the impulse relay time constant is set equal to the PI controller's reset time.¹ The theory is discussed briefly in Chapter 12.

9.7 OVERRIDES FOR COLUMN OVERHEAD SYSTEM

Let us assume we have a conventional column with the following normal controls:

- Condensate receiver level controls top-product drawoff.
- Base level controls bottom-product drawoff.
- Reflux is ratioed to feed.
- Steam is ratioed to feed.

As indicated by Figure 9.9, this column has a horizontal condenser and vertical, cylindrical condensate receiver or reflux drum. We will assume that the level controller is of the PI type with set point at midscale of the level transmitter span and that gain 2 auto overrides are employed. The level overrides then function as described in the following.

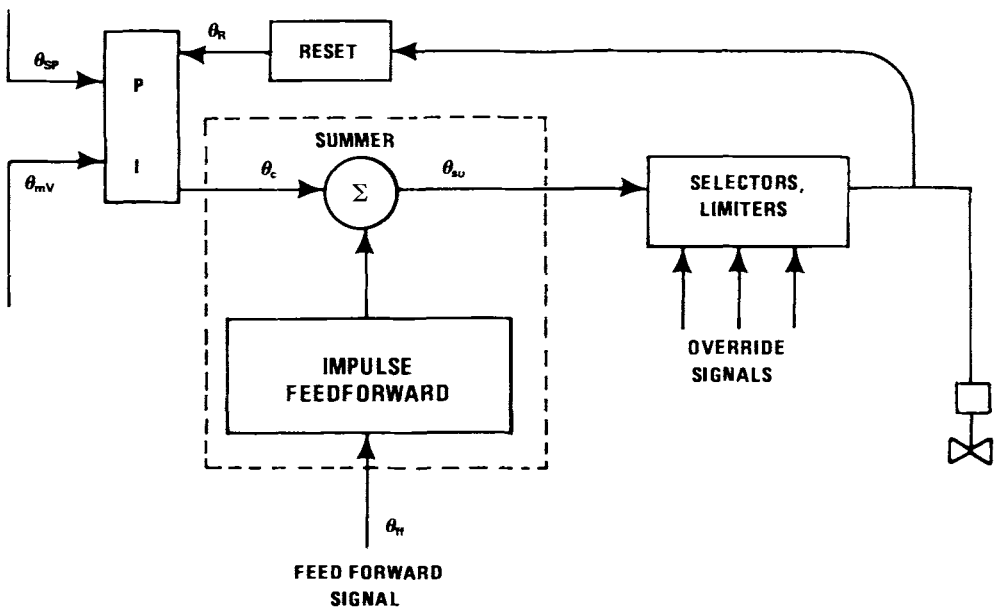


FIGURE 9.8
Impulse feed forward with PI controller and overrides

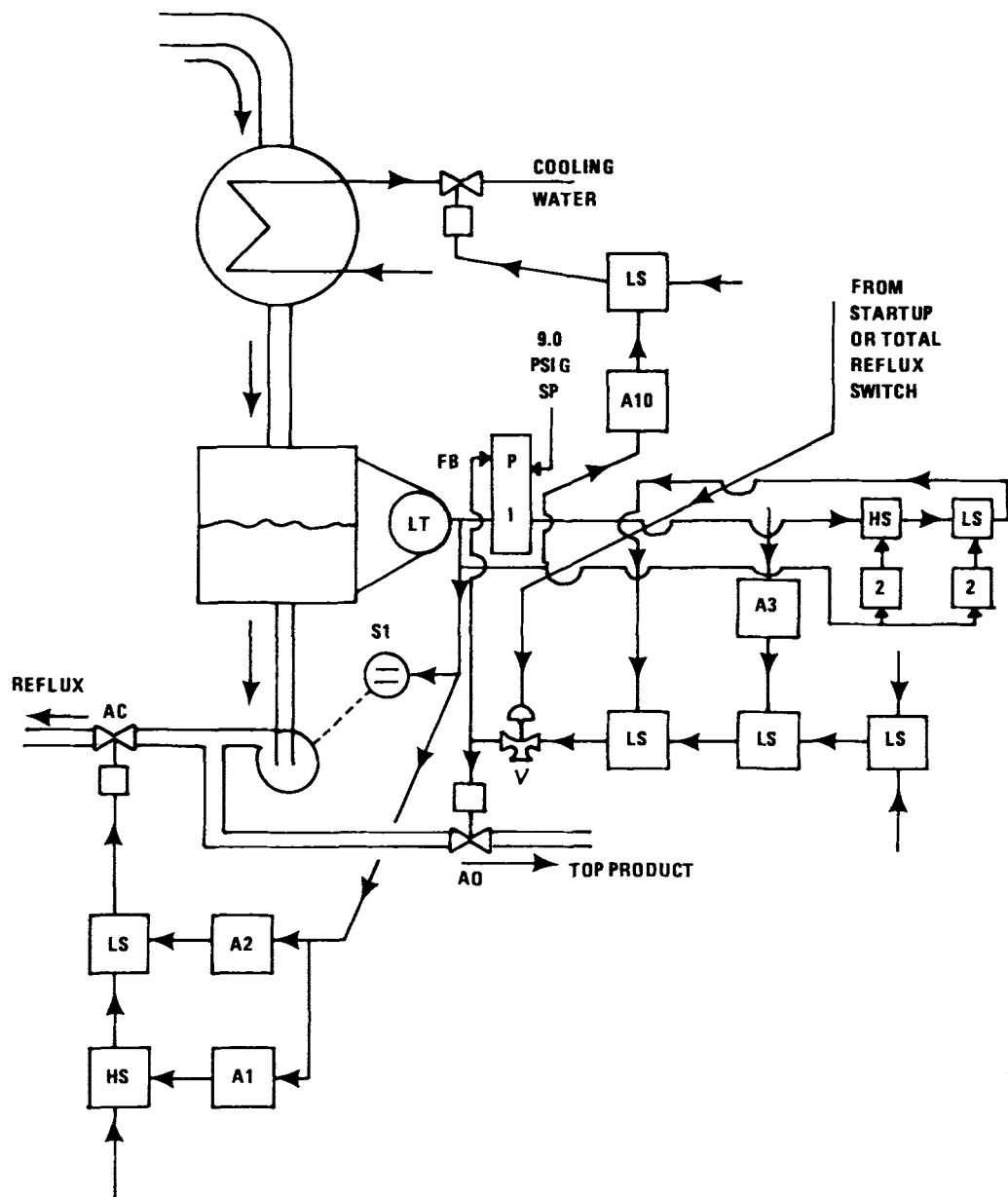


FIGURE 9.9
Overrides for column overhead system

Low Condensate Receiver Level Override on Reflux

As liquid level rises to the lowest position (3 psig), the output of reverse-acting override A1 with gain 4 starts to decrease from 15 psig; when the level reaches 25 percent (6 psig), the output of A1 is 3 psig. This permits the reflux valve to open and the reflux controller to take over somewhere in the lower 25 percent of level. If the level drops again, this override starts to close the reflux valve.

This design should be compared with that of Figure 3.26 where a PI level controller was used with auto overrides, and high and low level overrides acted on the reflux flow controller set point. This permits accurate calculation and prediction of override behavior, and overall is probably preferable. We are using this scheme increasingly. It has been a more common practice in the past, however, to have overrides act directly on the valve as in Figure 9.9.

High Condensate Receiver Level Override on Reflux

If liquid level rises above 75 percent, the output of reverse-acting override A2 with gain 4 starts to decrease from 15 psig and reaches 3 psig at 100 percent level. This drives the reflux valve open. These two overrides are particularly helpful for a column with a high reflux-to-distillate ratio. Without such protection moderate changes in reflux flow could either flood the receiver or run it dry.

Low-Low Condensate Receiver Level Override on Cooling Water

To guarantee condenser cooling-water flow during startup, override A10 holds the cooling-water valve open as long as the condensate receiver level is low. A10 usually has a gain of -25 and is so calibrated that it has an output of 9 psig when its input is about 3.5 psig. The normal controller for this valve may be the cooling-water exit-temperature controller, or in the case of vacuum or pressurized columns, it may be the pressure controller.

Overrides on Distillate Valve

The distillate valve may be held closed by override A3 until overhead composition or temperature reaches a satisfactory value; it may also be closed by high liquid level in the next process step. Reflux pump operation is made automatic by switch S1. As the condensate receiver level transmitter output increases above 3 psig, S1 starts the pump. Similarly, if the reflux drum loses its level, the pump shuts down.

For some columns it is desirable to maintain a minimum reflux flow rate. This would require an additional override on the reflux valve.

Condenser-Cooling-Water Overrides

As mentioned in Chapter 3, some plants like to control condensate temperature by throttling cooling water as shown in Figure 3.2. This seldom works well,

if at all, so it is increasingly common to provide an override from cooling-water exit temperature as shown by Figure 9.10. If the condensate temperature becomes too low, causing the condensate temperature controller to cut back on cooling water, the cooling-water exit temperature rises. To minimize fouling and corrosion, it is usually desirable to limit exit cooling-water temperature to a maximum of 50–60°C, particularly if 316 SS is used. The override scheme of Figure 9.10 will accomplish this.

9.8 OVERRIDES FOR COLUMN-BASE SYSTEM

The column-base system shown by Figure 9.11 features a thermosyphon, vertical-tube reboiler. The base liquid level transmitter is installed and calibrated

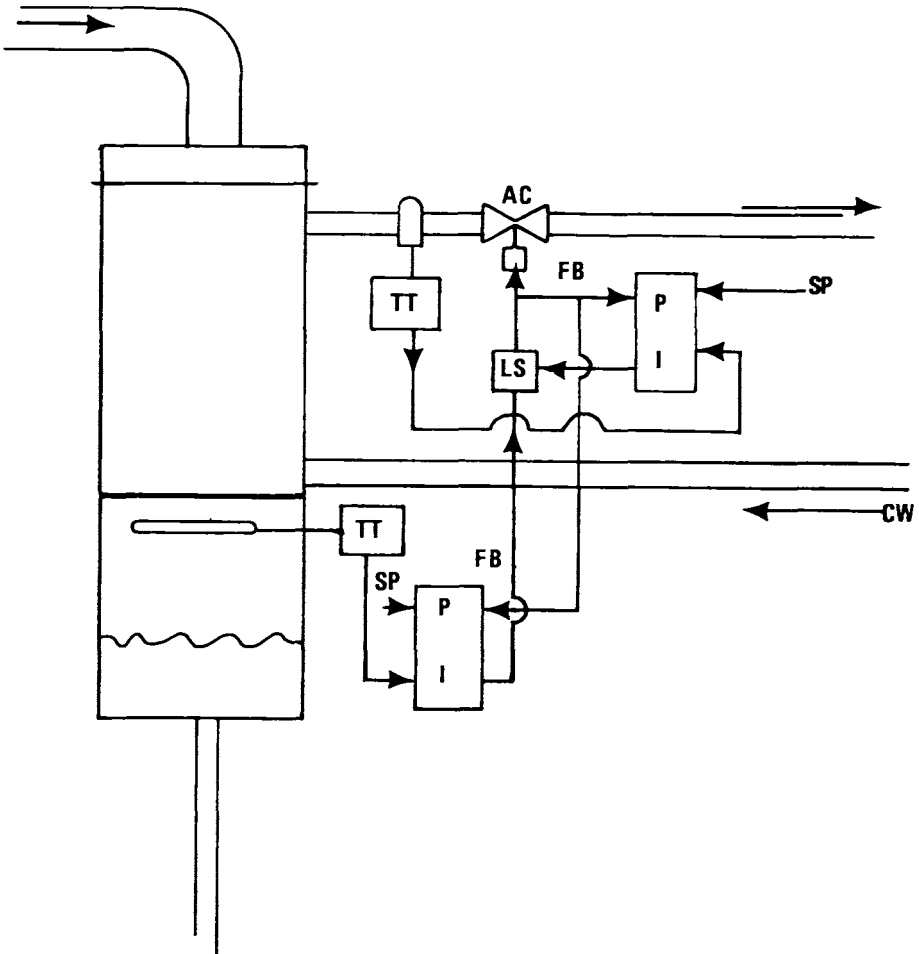


FIGURE 9.10
High cooling water exit temperature override on condensate temperature control

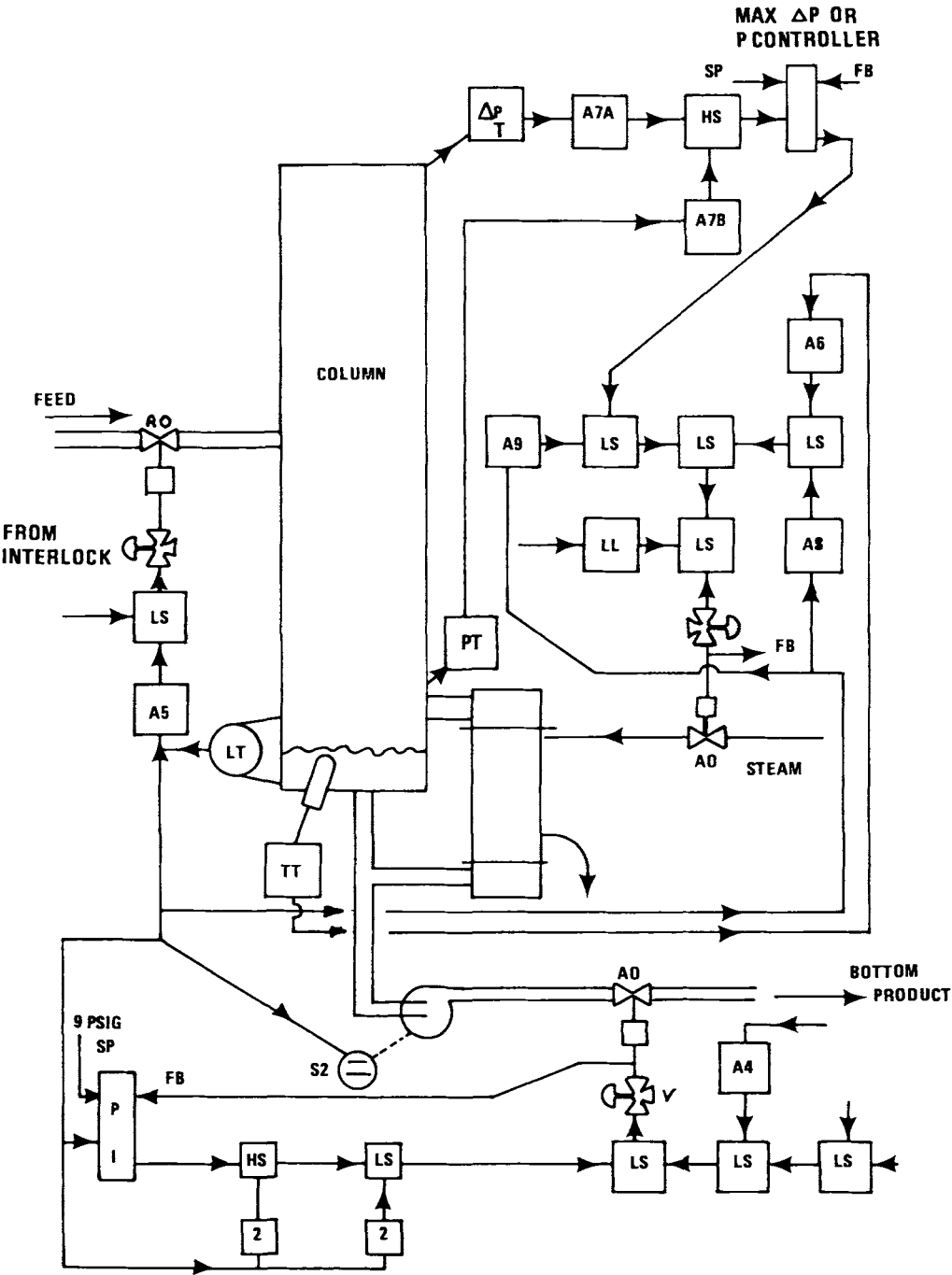


FIGURE 9.11
Overrides for column base system

as recommended in Chapter 4. The PI controller has gain 2 auto overrides, and the set point is at midscale. The level overrides then function as in the following.

High-Base-Level Override on Feed

As liquid level in the base rises from 75 to 100 percent, the output of gain -4 override A5 goes from 15 to 3 psig; this closes the feed valve through a low selector.

Low-Level Override on Steam

As liquid level rises from zero to 25 percent, override A9 output increases from 3 to 15 psig. This permits the steam valve to open and the steam controller to take over somewhere in the lower 25 percent of level. If the level drops below 25 percent, the override starts to close the steam valve. This prevents "baking" the reboiler tubes (which fouls them) when there is too little liquid to keep the tubes wet.

High-Base-Pressure or High-Column ΔP Override on Steam

As increasing boilup increases column ΔP or column-base pressure (especially during startup), override A7A or A7B, acting through a proportional-reset controller, will limit maximum steam flow. This is accomplished by setting a 9-psig set point on the controller and biasing gain 1 relays A7A and A7B such that their outputs are each 9 psig when the maximum permissible column ΔP and base pressure are reached. Whichever is reached first will then initiate override action.

A convenient way of determining maximum column ΔP is from the equation:

$$\Delta P_{\max} = 1.2 \Delta P_{FS} \quad (9.5)$$

where ΔP_{FS} is the flowsheet value of ΔP . This is based on the common practice of designing columns such that the flowsheet value of boilup is about 80 percent of the boilup that would cause flooding. The factor of 1.2 gives a ΔP_{\max} just short of flooding. For determining maximum permissible pressures for overrides and interlocks, consider four pressures:

- Relief valve set pressure, P_{RV}
- Interlock set pressure, P_{IN}
- Override set pressure, P_{OR}
- Normal operating pressure, P_N

A convenient way of setting P_{IN} and P_{OR} is to let

$$P_{IN} = P_N + 1/2 (P_{RV} - P_N) \quad (9.6)$$

and

$$P_{OR} = P_N + 1/4 (P_{RV} - P_N) \quad (9.7)$$

Minimum Steam Flow Controller

As liquid level rises from 75 to 100 percent, override A5 closes the feed valve. At a level transmitter output of 15 psig, A5 has an output of 3 psig. With steam/feed ratio control, this would also cut off steam. But it is desirable to maintain enough boilup that the trays (sieve or valve) do not dump or weep. For this we provide an override (not shown) that is a minimum steam flow controller. In most cases we have found it advisable to adjust override biases so that a high level pinches feed before steam.

Miscellaneous Column-Base Overrides

The base system contains two more overrides. During startup A4 holds the tails valve closed until base composition reaches a desired value. A6 closes the steam valve if base temperature becomes too high.

It should also be noted that the bottom-product valve may be closed by an override from the next process step. The bottom-product pump is operated automatically by switch S2, which turns the pump on whenever the base level transmitter increases above 6 psig.

9.9 AUTOMATIC STARTUP AND SHUTDOWN

Two kinds of startup and shutdown are commonly encountered, depending on whether the column base is full ("wet") or empty ("dry") when the column is shut down. As will be seen, wet column startups and shutdowns are much faster. The same basic override system permits automatic startups regardless of whether they are wet or dry. It should be noted that no formal program or sequence control is necessary. Instead, as soon as steam and feed are available, the column comes on line at the maximum speed permitted by the constraints. Depending on circumstances, a column may or may not follow the same sequence of constraints from startup to startup.

Wet Column Startup

If a column is shut down by closing the steam and drawoff valves, the entire column contents accumulate in the lower section of the column. Typically the liquid level rises well above the vapor inlet from the reboiler and up over some of the lower trays. (Some column designers prefer to provide enough space below the first tray that it can never be flooded.) Such a shutdown may be made to accommodate production scheduling or may have been forced by process interruptions elsewhere. The column may be started up again simply by turning on the steam.

As shown in Figure 9.6, an automatic startup may be accomplished with an air switch connected to a three-way, air-operated valve in the air line between the steam valve and its overrides. The "on" position corresponds to a through

connection; in the “off” position, the output of the overrides is dead-ended and the signal connection to the valve is vented to the atmosphere. It is probably preferable for each column to have its own switch, but a master switch can be used to start a complete train.

The adjustable lag box (commonly an inverse derivative unit, or, in electronics a first-order lag or integrator) is so hooked up that when the switch is positioned to “operate,” the steam valve opens slowly. When the switch is moved to “shutdown,” the steam valve is closed immediately.

With all controls on automatic, including condenser cooling-water controls not shown in Figure 9.6, the operator pushes the switch to “on.” The steam valve immediately begins to open slowly, and continues opening until the column ΔP override pinches it a little. It stays at a partly open position until dropping base level closes it further. The column feed valve stays closed until falling base level permits it to open. Eventually decreasing low boiler concentration in the base permits the bottom-product valve to open and be controlled by base level. Column-base composition is then controlled by throttling the steam valve.

In the meantime inventory is starting to build up in the reflux drum. The column stays on total reflux until overhead composition is satisfactory and reflux drum level has risen high enough for the overhead composition controller to take over the reflux valve. Since this procedure permits maximum possible steam flow with only one or two constraints, it is the fastest possible procedure for getting a column on line. In addition, since each column in the train starts with adequate initial inventory, it is not necessary for downstream columns to wait for upstream columns to get started. This procedure, therefore, puts a train on stream much faster than does a dry column startup.

To shut down a column or train, it is only necessary to switch the master switch to “shutdown”; this shuts off the steam, feed, and drawoff valves. Liquid then drains down into each column base. This procedure has worked well for small columns, say, 6 feet or less in diameter. For larger columns there have been some cases where the lower trays were damaged. It is probably safer, therefore, before startup to pump column-base contents down to a level well below that of the bottom tray. In doing this one should also allow for displacement of liquid from a thermosyphon reboiler; once boilup has been established, most of its tube volume will be vapor.

Some plants have spare or utility tanks that can be used as temporary reservoirs for excess column-base contents.

Dry Column Startup

In this case, the columns have little or no inventory, top or bottom. Startup of the train is accomplished by turning the master switch to the “operate” position. As before all controls are initially on “automatic.” The steam valve starts to open as soon as base inventory builds up. All controls then operate as indicated previously. Since, however, the second column cannot start until

it begins to get feed from the first, it may take a long time to get the whole train going.

Shutdown is accomplished by shutting off feed to the first column. Each column then automatically works its overhead and base inventories down to the lower 25 percent zone. At this point the operator must exercise manual control of the bottom product and steam valves to work column inventories down to very low levels. When this has been done, the columns may be shut down by switching the master switch to "shutdown."

To get everything out of the columns, as might be required for maintenance work, manual drain valves must be opened and residual column contents taken off in drums or other containers.

9.10 "IDLE" OR TOTAL REFLUX

The controls discussed are capable of operating the columns in an "idling" or total reflux condition with either a high or a low column inventory.

High Column Inventory

If, during normal operation, the bottom- and top-product valves of a column are closed manually or are held closed by a total reflux switch, the overhead and base inventories build up until the feed valve is shut off. The column is then on total reflux with the steam valve controlled by ΔP and reflux flow set by condensate receiver level.

The column can be put back on normal control by switching the two level controls to automatic, or by the total reflux switch.

Low Column Inventory

If, during normal operation, the feed valve to a column is closed manually or closed by some upstream override, the column will work its overhead and base inventories down to the 25 percent levels and then put itself on total reflux. Normal operation is restored by putting the feed valve back on automatic, or by coming out of the override condition. Note that feed pump failure will put the column on total reflux by this method.

For total reflux a separate switch may be provided that actuates air-operated, three-way valves in the air lines to the drawoff valves. The startup/shutdown switch then controls only the feed and steam valves. This design permits the operator to put the column into total reflux, when desired, or to take it out of total reflux during startup upon ascertaining that it is proper and safe to do so.

Product Recycle to Feed Tank

To minimize the transition from total reflux to normal operation, it has been proposed to switch first from total reflux to recycling product streams

back to the feed tank. Then when the column is lined out—that is, making the desired separation—one may close the lines back to the feed and open the normal product lines to storage or the next process step. This eliminates the upset that sometimes occurs when switching from total reflux to normal operation. We are not aware, however, of any industrial applications.

9.11 MISCELLANEOUS OVERRIDES

Centrifugal Pump Bypass

At startup and shutdown, it is sometimes necessary to have a centrifugal pump running for a period of time while no product is being taken. It is undesirable, however, to “dead head” a centrifugal pump very long. Various expedients have been devised to take care of this. One popular method is to provide a bypass line with a restricting orifice. This is a low-investment design but it wastes horsepower.

Another approach, shown in Figure 9.12, uses a small valve in a bypass line. The controls are so arranged that as the flow approaches the specified

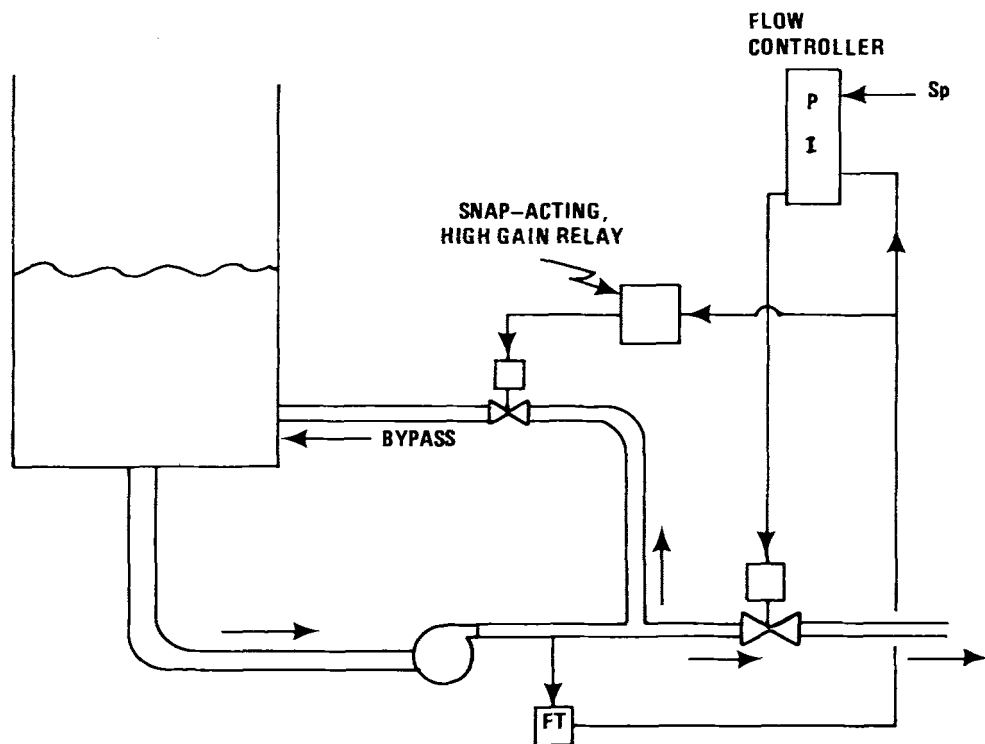


FIGURE 9.12
Scheme for protecting centrifugal pump against dead heading

minimum value, a high-gain relay opens the normally closed bypass valve. This is not an override in the normal sense since no signal selectors are needed. The high-gain relay should have hysteresis or detent to prevent chatter.

If the bypass line cannot be connected back to a tank or receiver as indicated, but must be connected to pump suction, a cooler should be installed in it to prevent overheating. If a flow measurement is not available, the high-gain relay can be set to open the bypass valve as the normal valve approaches the closed position.

Overrides for Maximum Capacity

Where it is desired to be able to run a column or train at maximum capacity, certain additional overrides may be needed.

A. Entrainment Override

With the previously discussed system, the limitation on any one column is the maximum permissible column ΔP that was taken to be a known, fixed value. For some columns, however, as one increases feed rate slowly and before flooding starts, a plot of top composition, x_D , versus reflux flow, w_R , shows a maximum value of x_D at a particular reflux rate as shown by Figure 9.13. To put it another way, at some high feed rate, an *increase* in reflux flow will cause a decrease in overhead purity, x_D . This is so because the increase in entrainment offsets the effect of increased reflux. This usually occurs at a column ΔP considerably higher than the nominal maximum permissible ΔP . It should be noted, however, that the peak of the x_D versus w_R curve does *not* occur at fixed, reproducible values of x_D , w_R , and ΔP . By measuring x_D and w_R and computing dx_D/dw_R , one can devise an override control system that maintains a positive value of dx_D/dw_R by pinching the steam or feed valve. This would replace the ΔP override. Implementation is shown in Figure 9.14.

B. Limited Utility Override

If the condenser or reboiler limits before the column proper does, one may use either the cooling-water valve position or steam-valve position for override purposes. Generally, for good control, it is not desirable for valves to go more than 95 percent open or less than 10 percent open, except on a momentary basis. Overrides may be provided such that if either valve goes beyond this value, the feed valve is pinched slowly until the utility control valve's position is back to a reasonable value. This requires a slow-acting override incorporating a proportional-reset or floating controller. A particular arrangement is shown in Figure 9.15.

For highly integrated plants, it may be necessary to protect the utility supply during startups. In "bootstrap" plants, for example, it may be necessary during wet startups to avoid putting such a demand on the steam header as to reduce header pressure seriously. One automatic approach is to connect header pressure to overrides on the reboiler steam valves to close them partially as header pressure drops. A possible hookup is shown on Figure 9.16.

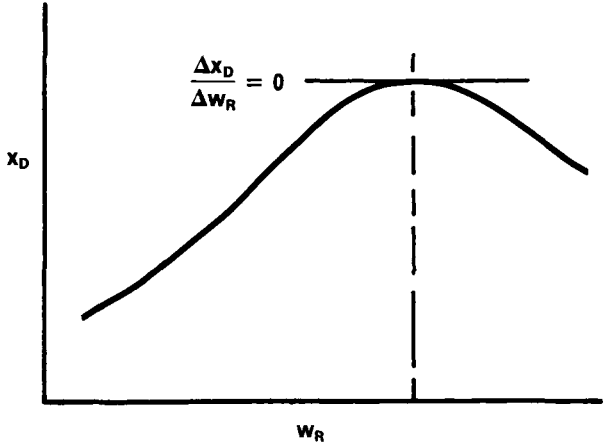


FIGURE 9.13
Effect of entrainment on overhead composition

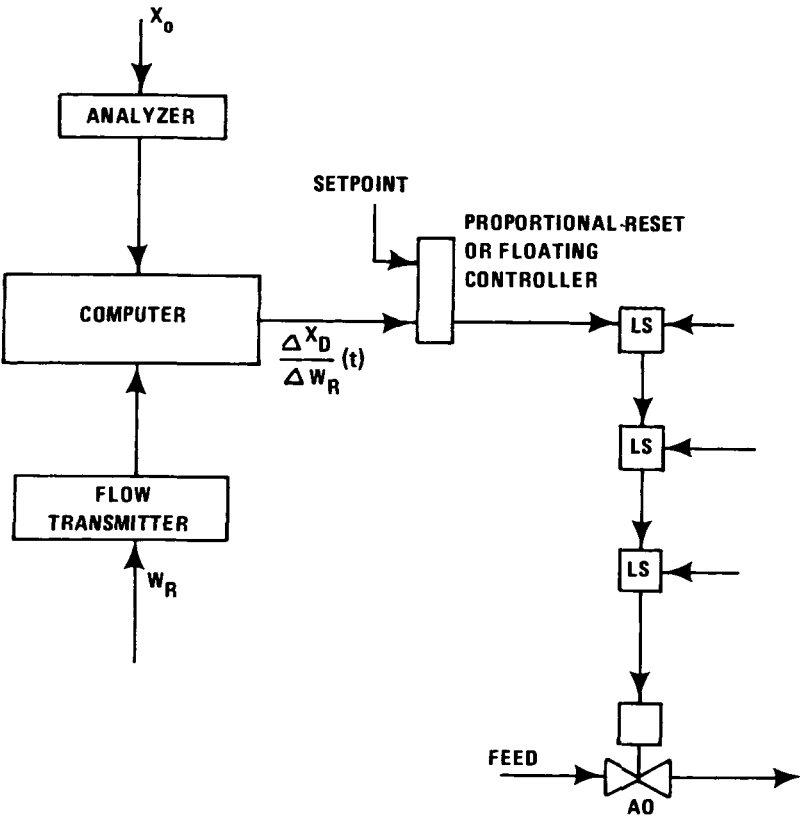


FIGURE 9.14
Entrainment override

C. Automatic Balancing of Condenser and Reboiler Heat Loads

If the condenser and reboiler do not have balanced heat-transfer capability, one of them will limit before the other does as feed rate increases, provided the column itself does not limit first. One way of balancing these two heat exchangers on pressurized or vacuum columns, thereby permitting maximum column capacity, is to feed the steam valve and cooling-water valve positions into a summer with gain whose output becomes the “set point” of the column pressure controller as shown in Figure 9.17. Note that a high limiter holds signal *A* constant for cooling-water-valve loading signals greater than 4.2 psig and a low limiter holds *B* constant for steam-valve loading signals less than 13.8 psig. The summer bias, *C*, then becomes the pressure controller normal set point.

If, for example, the condenser cooling-water valve opens wider than the steam valve, and if its loading signal is less than 4.2 psig, the summer or “balance” controller calls for a slightly higher column pressure to increase condenser heat transfer.

The column and auxiliaries, of course, must have adequate static pressure ratings, and the required range in pressure must not be so large as to cause adverse changes in relative volatility.

9.12 DESIGN CONSIDERATIONS

It is probably apparent that most overrides are really feedback control loops. They therefore are subject to stability considerations. In many cases they are also subject to truly hard constraints, as, for example, maximum column-base pressure. Since any feedback control system must have some room within which to work, the overrides must be so designed that the process does not normally approach the hard constraints too closely.

In the case of overrides with proportional-only action, we can visualize, as shown in Figure 9.18, a zone between hard and soft constraints. The width* of this zone is determined by the override loop gain, which is limited by stability considerations. With a proportional control loop designed for dead-beat response, there is a unique relationship between the value of the manipulated variable and the distance between the process variable and its hard constraint. The manipulated variable always reaches its maximum (or minimum) value before the process variable exceeds its hard constraint. The soft constraint will correspond to the minimum (or maximum) value of the override output. The takeover point between normal and override controls will be at a variable position (depending on operating conditions) somewhere between the hard and soft constraints.

* It is very helpful to think of this zone as a fraction of the measured variable transmitter span.

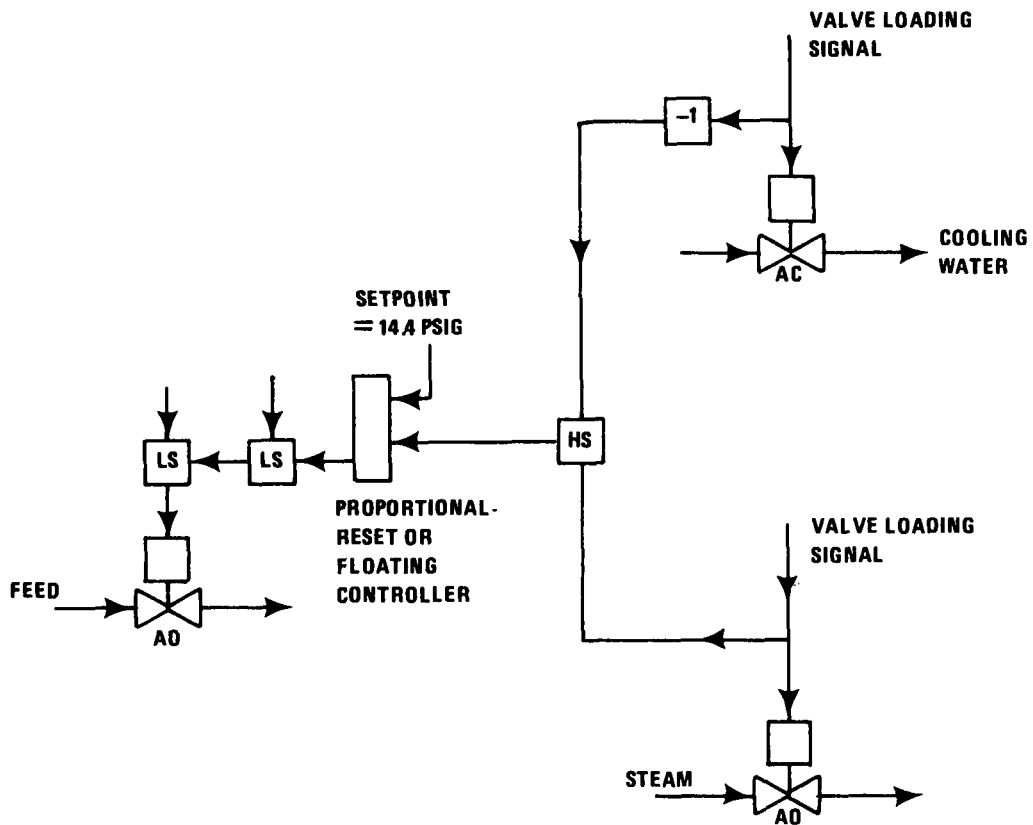


FIGURE 9.15
Limited utility override on feed

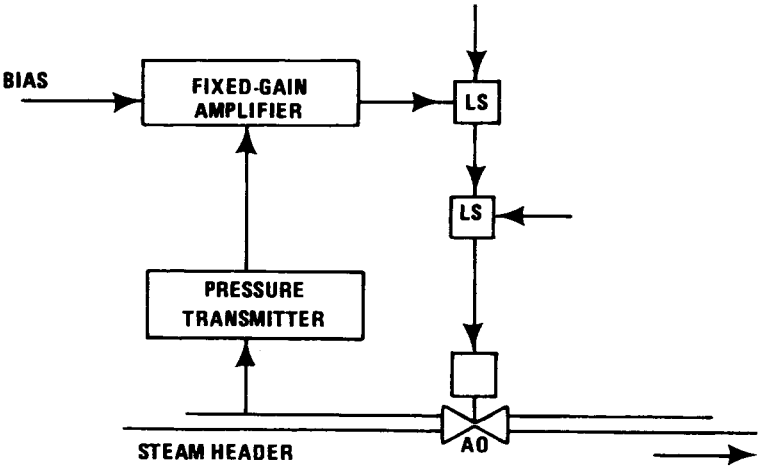


FIGURE 9.16
Steam header pressure protective override

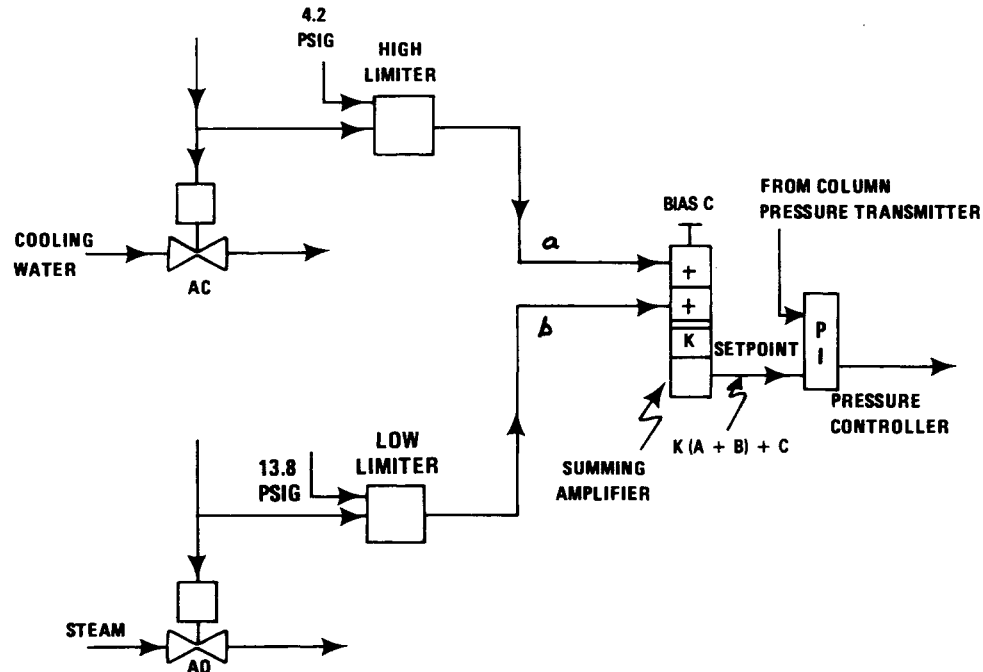


FIGURE 9.17
Control scheme for balancing condenser and reboiler heat loads

9.13 OVERRIDES FOR SIDE-DRAW COLUMNS

We will conclude this chapter with a brief discussion of some of the overrides that may be encountered on side-draw columns. We will choose, as an example, the column of Figure 9.19. This is a purification column in a solvent-recovery system. A small amount of low boilers must be removed overhead, and a small amount of high boilers must be removed as bottom product. Most of the feed is taken off as sidestream. Following are the most important overrides.⁴

1. High and low feed-flow limiters are set according to the required column turndown (ratio of maximum specified feed flow to minimum specified feed flow); see Figure 9.20. Increasingly we are putting the limiters in the set-point channel instead of as shown.

2. Low base temperature or low top temperature holds all drawoff valves closed (Figure 9.21). Usually a gain 6 override is adequate, but this really depends on column composition dynamics. The function here is to hold the column on total reflux until overhead and base compositions are nearly correct. On-stream analyzers would permit a better job here.

3. High column ΔP or high base pressure closes the steam valve. A proportional-reset controller is usually necessary here. Typical gain is 0.5 and reset time is usually 20–40 seconds. Since column dynamics are about the same for either variable, only one controller is necessary if the two transmitters are scaled to have the same gain. A high selector chooses between them. All overrides for the steam valve are shown in Figure 9.22.

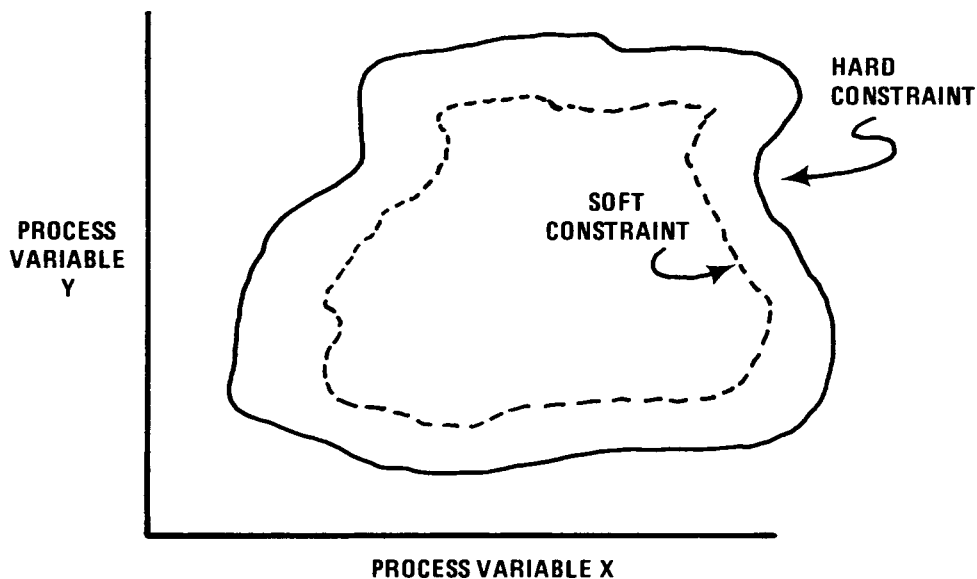


FIGURE 9.18
Hard and soft constraints

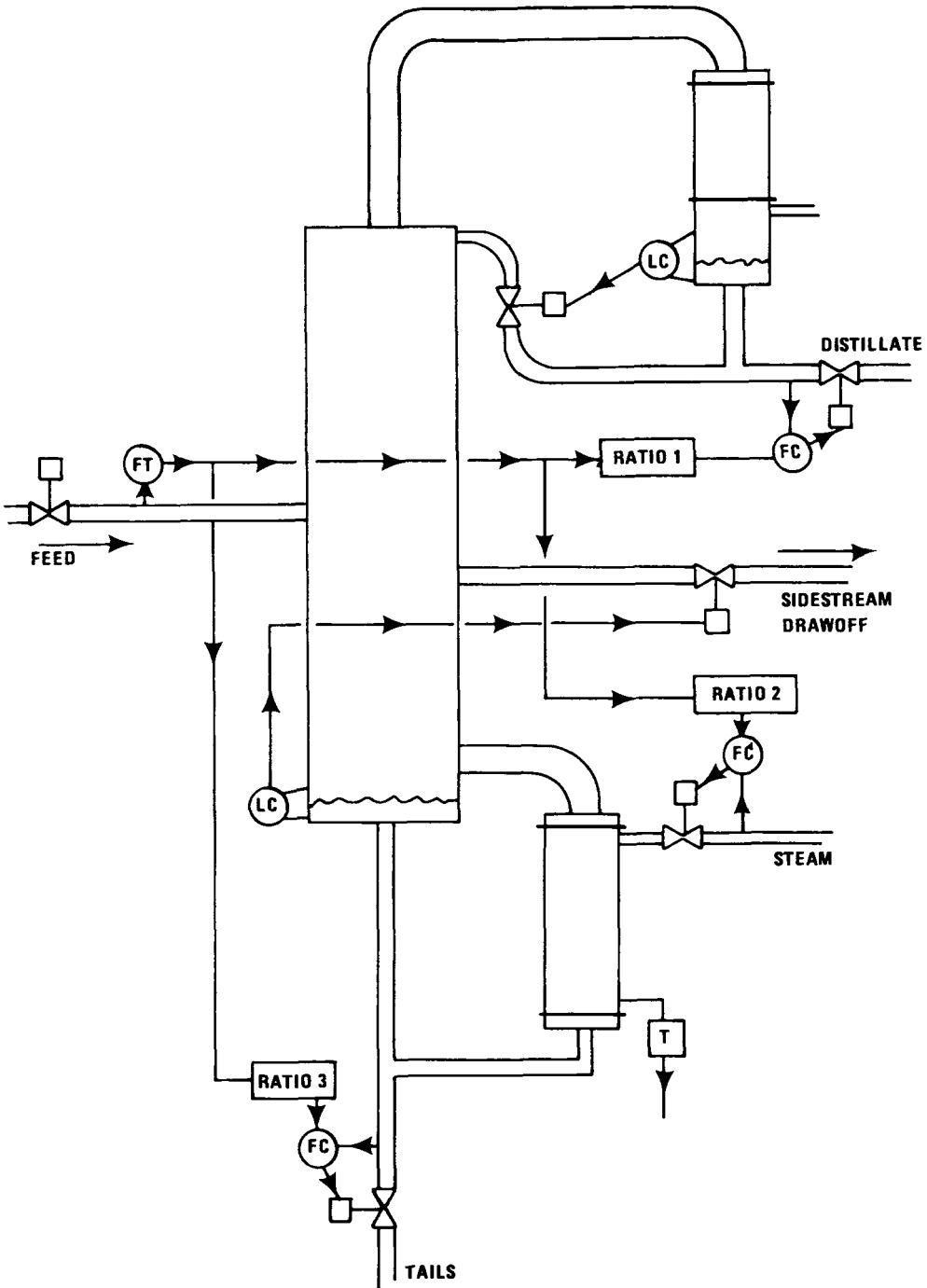


FIGURE 9.19
Flow rate controls for composition control

4. For columns with a sidestream *vapor* drawoff, it is necessary to maintain a minimum vapor flow up the column *above* the drawoff point. As shown by Figure 9.23, this is accomplished by calculating the total vapor flow from the reboiler from the steam flow, and subtracting the sidestream vapor flow to get net vapor flow up the column. This signal becomes the measured variable to a proportional-reset controller whose set point is the minimum vapor flow called for by the column designer. The output from this controller goes to a low selector in the path between the base level controller and the sidestream drawoff valve. If, then, calculated vapor flow up the column becomes less than required, the override controller closes the sidestream drawoff valve just enough to force the required additional vapor up the column.

5. For columns with a sidestream *liquid* drawoff, it is necessary to maintain a minimum liquid flow down the column *below* the drawoff point. As shown by Figure 9.24, this is accomplished by subtracting the sidestream drawoff flow from the estimated internal reflux to calculate the net liquid flow down the column. If this flow is insufficient, the override controller pinches back on the drawoff valve until the downflow becomes adequate.

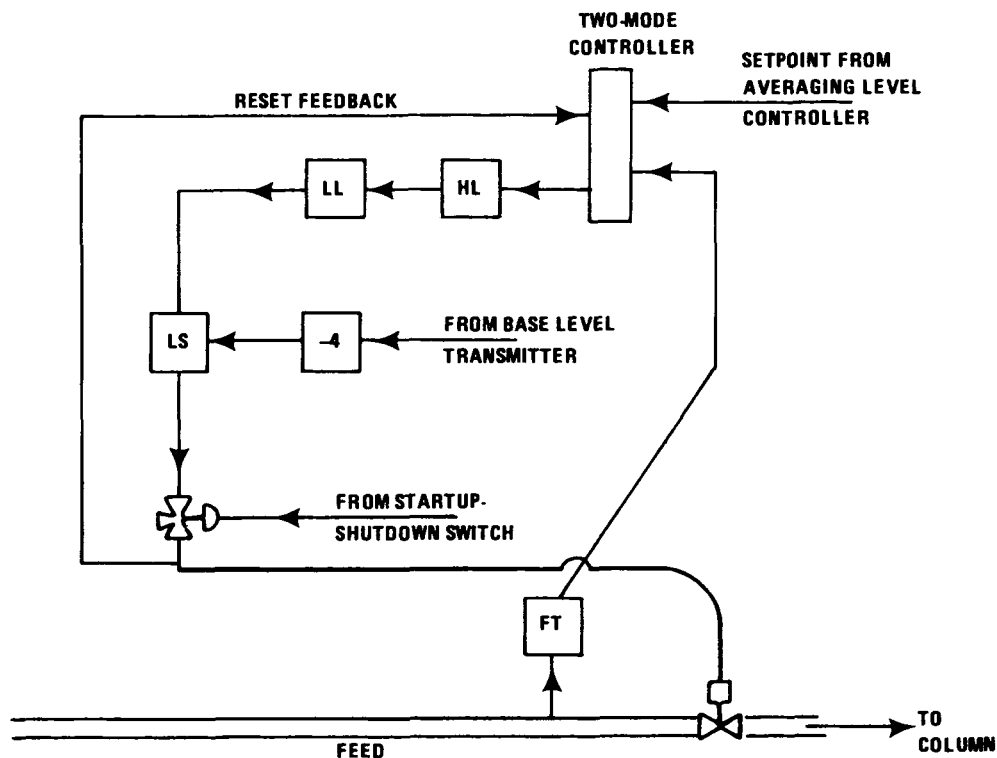


FIGURE 9.20
Feed flow system

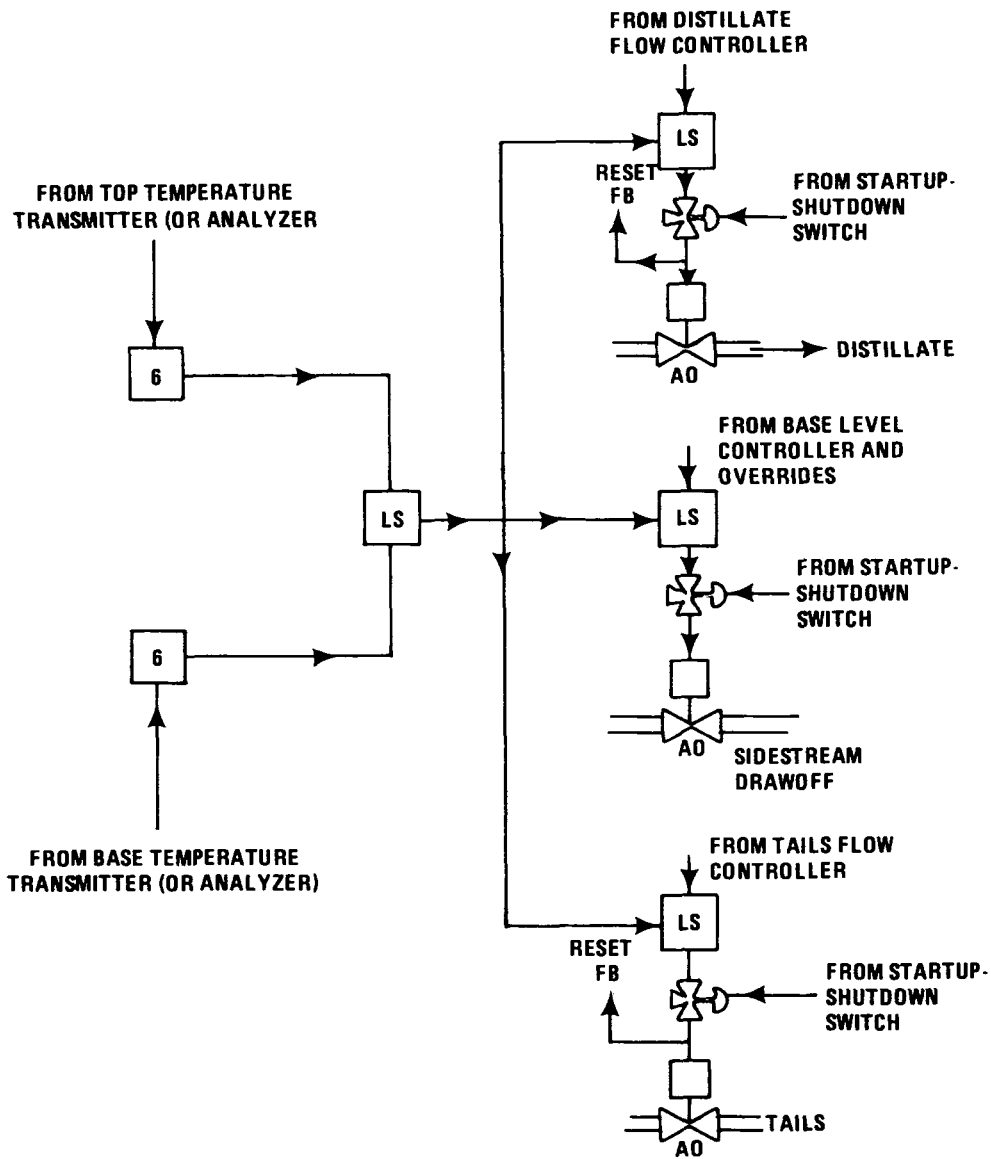


FIGURE 9.21
Low temperature overrides for drawoff valves

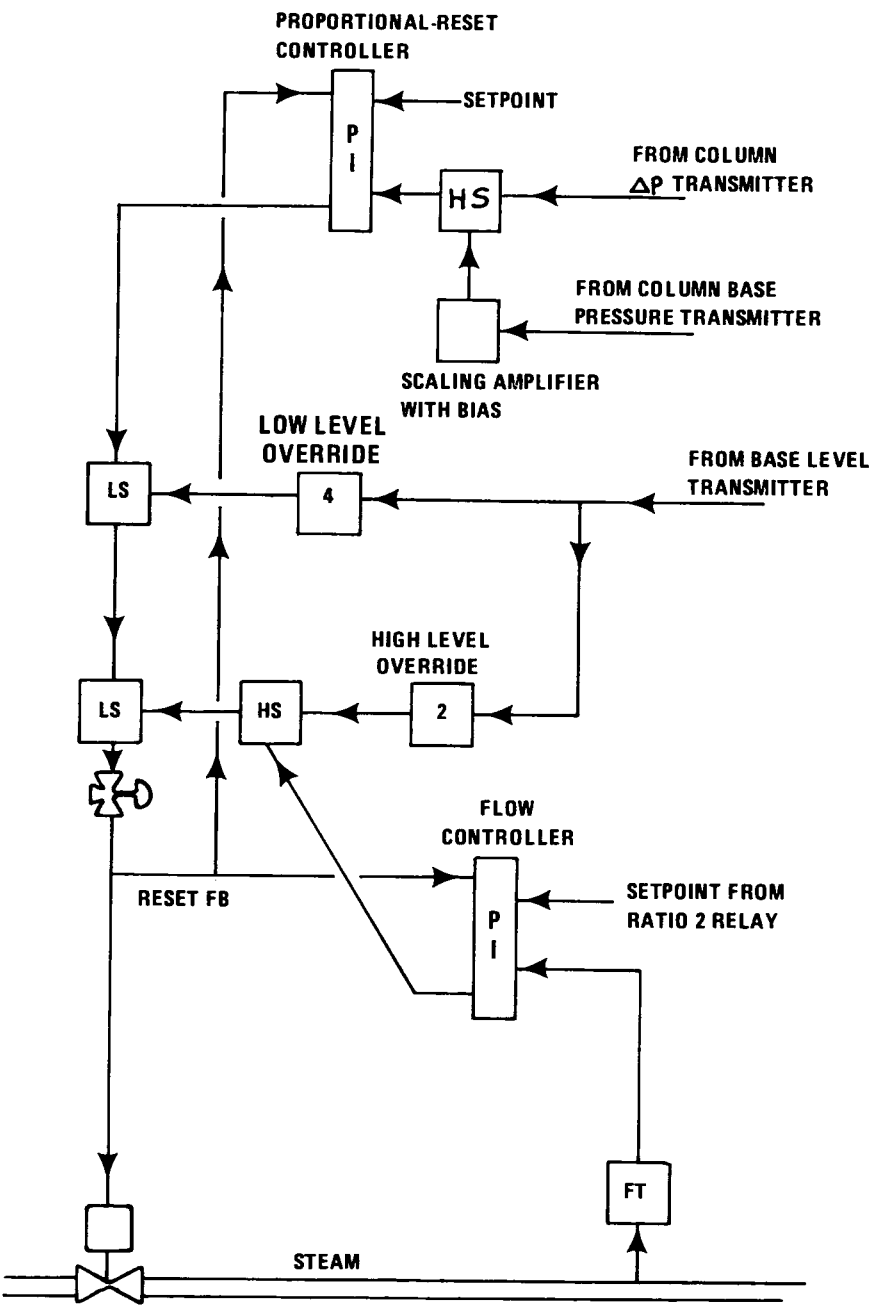


FIGURE 9.22
Steam valve overrides

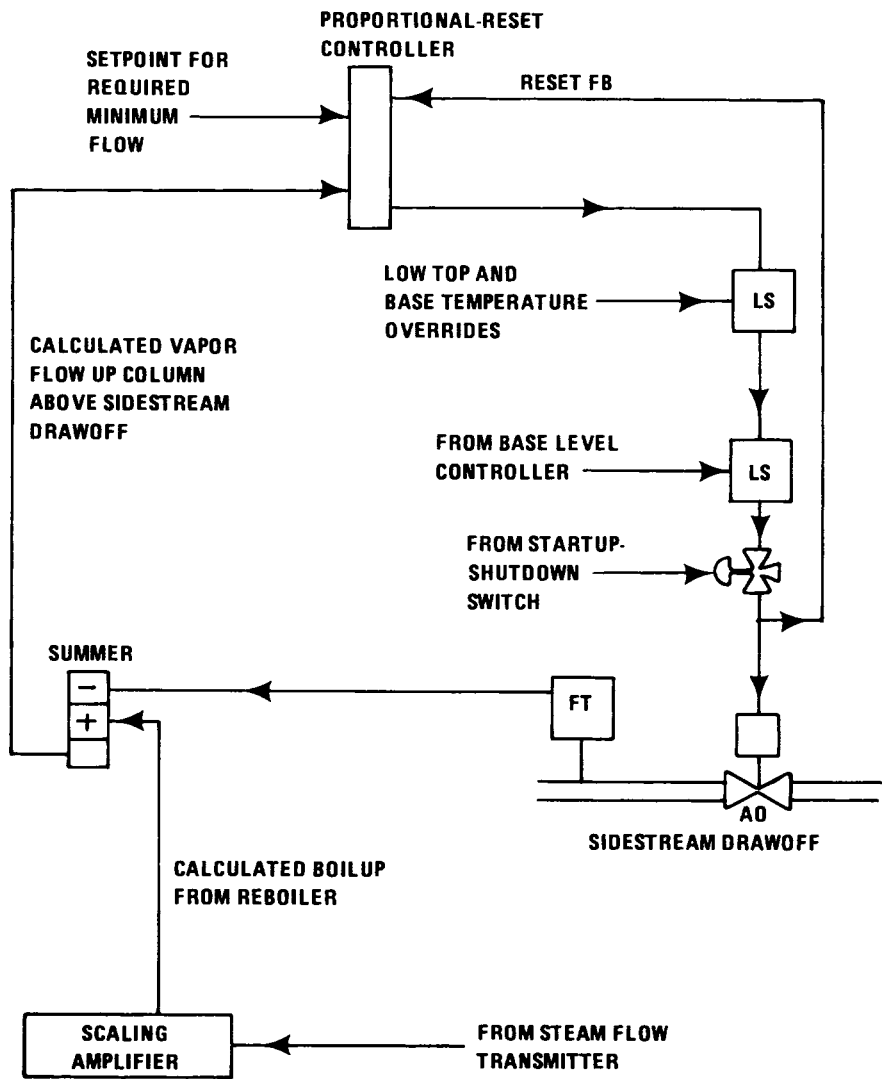


FIGURE 9.23
Override for minimum vapor flow up column

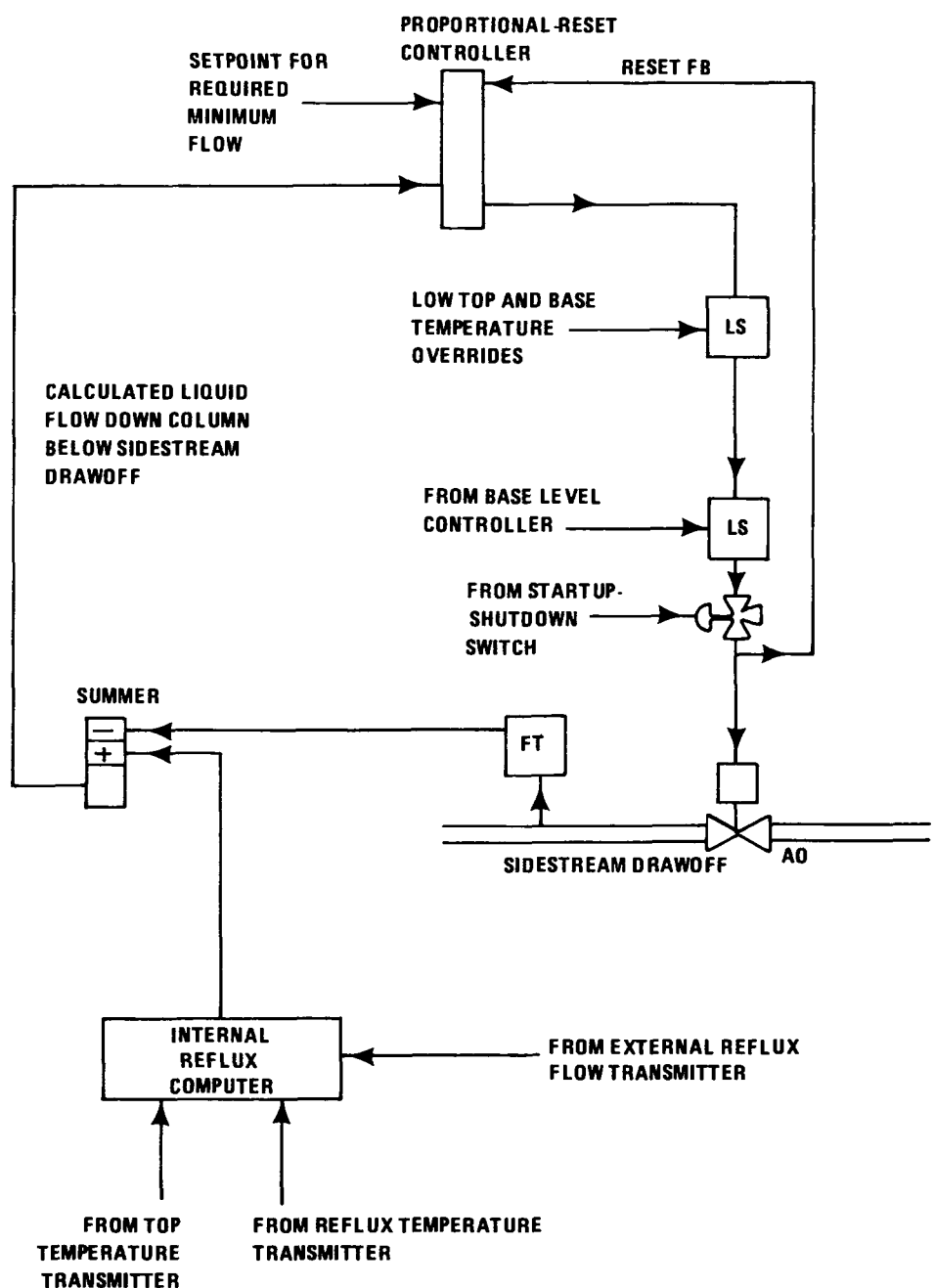


FIGURE 9.24
Override for minimum liquid flow down column

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10

Indirect Composition Measurements

10.1 INTRODUCTION

in recent years on-line analyzers have vastly improved in sensitivity, selectivity, speed of response, stability, and reliability. Consequently it is increasingly feasible to measure the composition of column product streams directly and use these measurements for column control.¹ But analyzers are expensive and are somewhat more trouble-prone than the simple devices used for measuring temperature, pressure, and flow. Consequently, particularly for less critical applications, considerable ingenuity has been expended on the use of these simpler instruments in various combinations to deduce compositions. For the most part, the techniques to be discussed have been demonstrated only for binary and almost-binary systems. For close control of multicomponent systems, analyzers are virtually mandatory.

This chapter discusses a number of indirect composition-estimation techniques. They vary from the measurement of a single temperature on a tray somewhere in the column to the measurement of a number of variables that are fed into an on-line computer that performs rigorous tray-to-tray calculations.

10.2 SINGLE-TRAY TEMPERATURE

Probably the most commonly used composition-estimation technique is the measurement of a temperature on a single tray in the column. This tray is normally located in the rectifying section if distillate composition is more important, or in the stripping section if bottom composition is more important. On some columns two temperatures are controlled, one in each section.

The selection of the location of the control tray has been the subject of numerous papers over the last 20 to 30 years. Rademaker *et al.*² present a well-

organized summary of the various ideas and criteria generated over the years. The normal procedure is to plot the steady-state temperature profile and then select a tray that is somewhere in the region where the temperature is changing fairly rapidly from tray to tray.

Theoretically, the temperature at the end of the column should be controlled in a binary constant-pressure system to maintain constant product composition. However, the temperature changes at the ends of the column are quite small in moderate- to high-purity columns. Therefore, small changes in pressure or the presence of other lighter or heavier components can affect temperature much more than composition of the key component.

Another consideration is nonlinearity. The relationship between tray temperature and the manipulated variable can be quite nonlinear if a tray near the end of the column is used. For example, suppose we have a system in which reflux is on flow control and a temperature near the bottom of the stripping section is controlled by heat input to the reboiler. An increase in heat input will drive more light components up the column, but the control tray temperature will increase only very slightly because it is essentially pure high boiler already. However, a decrease in heat input will drop light component down in the column and the control tray temperature can change very drastically. This nonlinear response presents difficult controller tuning problems when conventional linear controllers are used. Thus a control tray should be selected that is as close to the end of the column as possible but not so close that it gives a highly nonlinear response. Nor should it be so close that it is too sensitive to changes in pressure and to light or heavy nonkey components in the feed.

For good speed of response, temperature probes should be installed in the active part of a tray, not in downcomers.

10.3 DIFFERENTIAL TEMPERATURE

For a binary distillation, if the pressure is fixed, a single temperature measurement will provide a reliable guide to composition. But pressure is often not very constant. Today it is sometimes deliberately allowed to float to minimize energy consumption (see Chapter 8). Even if one fixes pressure at one end of a column, it will vary at the other end as a function of boilup rate. An early approach to compensating for pressure variations was to use two temperature measurements, one usually near one end of the column and the other at an intermediate tray. One was then subtracted from the other. For binary or almost-binary distillations, this works fairly well as long as boilup does not change much. If, however, boilup does change significantly, and if the two temperature measurements are separated by a substantial number of trays, one may encounter a nonmonotonic relationship between boilup and ΔT . As pointed out by Boyd,³ an increase in boilup tends to decrease ΔT due to increased purity but to increase ΔT due to increased pressure drop. The composite effect is that ΔT has a minimum at a particular boilup. This can be very confusing to the operator

and may cause instability in the ΔT control loop, and may either flood the column or shut it down.

This nonmonotonic relationship between ΔT and product composition occurs even when boilup changes do not affect ΔP . This is easily visualized if one considers two extreme situations. If the profile is so high in the column that the section between the two temperature points is filled with high boiler, the ΔT signal will be small. The ΔT signal will also be small if the profile is so low in the column that it is filled with low boilers. These two conditions will give similar ΔT signals but vastly different product compositions.

Efforts inevitably were made to stretch the ΔT technique to fit multicomponent distillations. Some of the difficulties encountered and some guidelines for successful application are presented by Webber.⁴

In another vintage paper, Vermilion⁵ presents a somewhat different application of differential temperature control. Here the intent apparently was not to compensate for pressure variations, but to control the temperature profile in the middle trays of the column, and thereby control terminal composition. For a 56-tray deisobutanizer, temperature was measured on trays 22 and 42 (measuring from top down). Use of the ΔT permitted successful control of isobutane in the bottom product.

Successful application to a C_3/C_4 splitter column was reported by Bonilla.⁶

10.4 DIFFERENTIAL VAPOR PRESSURE

Another early effort to compensate for pressure variations in binary distillations resulted in the development of the differential vapor pressure cell.⁷ The version made by Foxboro is known as the "DVP Cell." As shown in Figure 10.1, a bulb filled with liquid whose composition is the same as that desired on a particular tray is installed on that tray. It is connected to one side of a differential-pressure transmitter. The other side of the ΔP transmitter is connected directly to the same tray. When the liquid on the tray has the same composition as the liquid in the bulb, pressure in the bulb will be the same as pressure on the tray. For that condition the ΔP transmitter is normally set to read midscale. Deviations in the tray composition from that in the bulb are then reflected by ΔP transmitter output signals above or below midscale.

Where applicable, this instrument is capable of great sensitivity. There are, however, some practical problems:

1. Liquid of the desired composition for filling the bulb may not be chemically stable over a long period of time, thus leading to measurement errors.
2. Bulb filling is a job for the factory; reusing an instrument originally bought for another job may not be easy.
3. Response to pressure changes is much faster than the response to composition changes, which causes erroneous readings and sometimes control problems. Some users, therefore, have installed snubbers in the pressure impulse line.

Although this instrument is primarily intended for binary systems, Tivy⁷ has shown that with a bit of patience and cleverness, it can be used on some multicomponent systems.

To determine the required span of the DVP Cell in inches of water corresponding to a desired composition span, it is necessary to write and solve a few equations. If the reference bulb has been properly filled with a liquid of known composition, then at a reference temperature T_R pressure in the bulb is:

$$P_R = x_R P_L \gamma_{LR} + (1 - x_R) P_H \gamma_{HR} \quad (10.1)$$

where

P_R = total pressure, atm abs, in bulb

x_R = mol fraction low boiler in bulb

P_L = vapor pressure, atm, of low boiler at T_R

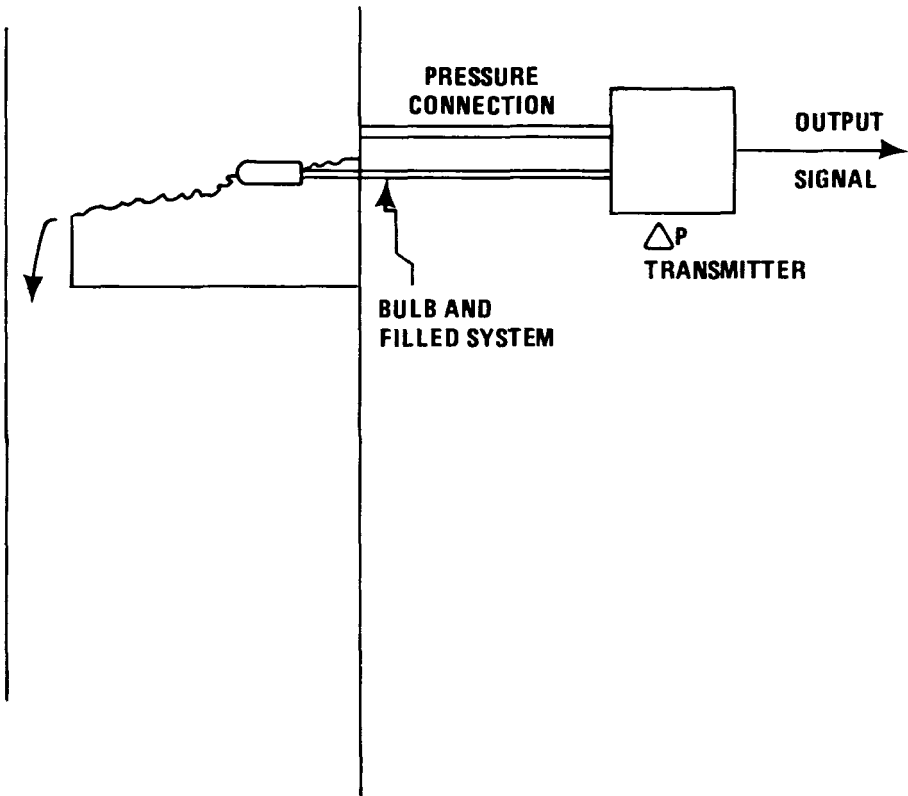


FIGURE 10.1
DVP cell schematic

P_H = vapor pressure, atm, of high boiler at T_R

γ_{LR} = low boiler activity coefficient in bulb

γ_{HR} = high boiler activity coefficient in bulb

At the same temperature, T_R , the process pressure is:

$$P_T = x_L P_L \gamma_L + (1 - x_L) P_H \gamma_H \quad (10.2)$$

where

P_T = total pressure in process, atm

x_L = mol fraction low boiler in process

P_L = vapor pressure, atm, of low boiler at T_R

P_H = vapor pressure, atm, of high boiler at T_R

γ_L = low boiler activity coefficient

γ_H = high boiler activity coefficient

Now the DVP Cell actually measures $P_T - P_R$ and is normally calibrated in inches of water for such spans as 20–0–20 or 50–0–50. From equations (10.1) and (10.2):

$$\begin{aligned} P_T - P_R &= x_L P_L \gamma_L + (1 - x_L) P_H \gamma_H \\ &\quad - [x_R P_L \gamma_{LR} + (1 - x_R) P_H \gamma_{HR}] \end{aligned} \quad (10.3)$$

which reduces to:

$$P_T = x_L P_L \gamma_L + (1 - x_L) P_H \gamma_H \quad (10.3a)$$

which is the same as equation (10.2).

We can now solve for x_L :

$$x_L = \frac{P_T - P_H \gamma_H}{P_L \gamma_L - P_H \gamma_H} \quad (10.3b)$$

Since P_R “washes out” of equation (10.3), one might well ask, “Why do we have it?” The answer is that there is only one instrument to calibrate rather than two, and the need for very accurately calibrated temperature and pressure measurements is eliminated.

The activity coefficients may be found from:

$$\gamma_L = \exp\left(\frac{A_{VL} (1 - x_L)^2}{[(A_{VL}/B_{VL}) x_L + 1 - x_L]^2}\right) \quad (10.4)$$

$$\gamma_H = \exp\left(\frac{B_{VL} x_L^2}{[x_L + (B_{VL}/A_{VL})(1 - x_L)]^2}\right) \quad (10.5)$$

where A_{VL} , B_{VL} are Van Laar coefficients. For other ways of finding activity coefficients, see the various texts on distillation and vapor-liquid equilibrium.

The vapor pressures of the pure components may be found by the Antoine equation:

$$P_L = e^{[A_L + B_L/(T + C_L)]} \quad (10.6)$$

$$P_H = e^{[A_H + B_H/(T + C_H)]} \quad (10.7)$$

where

A_L, B_L, C_L = Antoine coefficients for the low boiler

A_H, B_H, C_H = Antoine coefficients for the high boiler

T = temperature, °C

For the two-parameter Antoine equation:

$$C_H = C_L = 273.16$$

If, as we have postulated, the bulb contains the same chemical species as the process, $P_T - P_R$ will always be zero when the process has the same composition as the bulb. If the chemical species are the same but the compositions are different, $P_T - P_R$ will vary somewhat with pressure (and, therefore, temperature). That is to say, for $x_L - x_R$ constant, $P_T - P_R$ will change slightly as P_T changes. To facilitate calculations, an HP-41C program has been written that is valid if the process and bulb contain the same chemical species.

For systems that are not strictly binary but where the nonkey components are present in either relatively constant or very small concentrations, the DVP Cell may often be used by calibrating its output against laboratory analyses. When the desired composition is such that the components would react with one another over a period of time, one may substitute another fill material (if one can be found) that has the same P_R as desired at T_R . In this event the equations are slightly changed.

10.5 PRESSURE-COMPENSATED TEMPERATURE

To get around or away from the shortcomings of the previous methods, composition can be computed from temperature and pressure measurements. One technique is to measure pressure deviations from flowsheet pressure, P_{FS} , calculate temperature changes at constant composition due to these pressure deviations, and subtract these temperature changes from the actual temperature. In effect, the actual temperature is compensated back to what it would be at flowsheet pressure. A particular hardware concept (see Figure 10.2) we have used assumes that the P - T curve at constant composition is a straight line for small deviations. If, then, pressure deviations from P_{FS} are large, there will be some error.

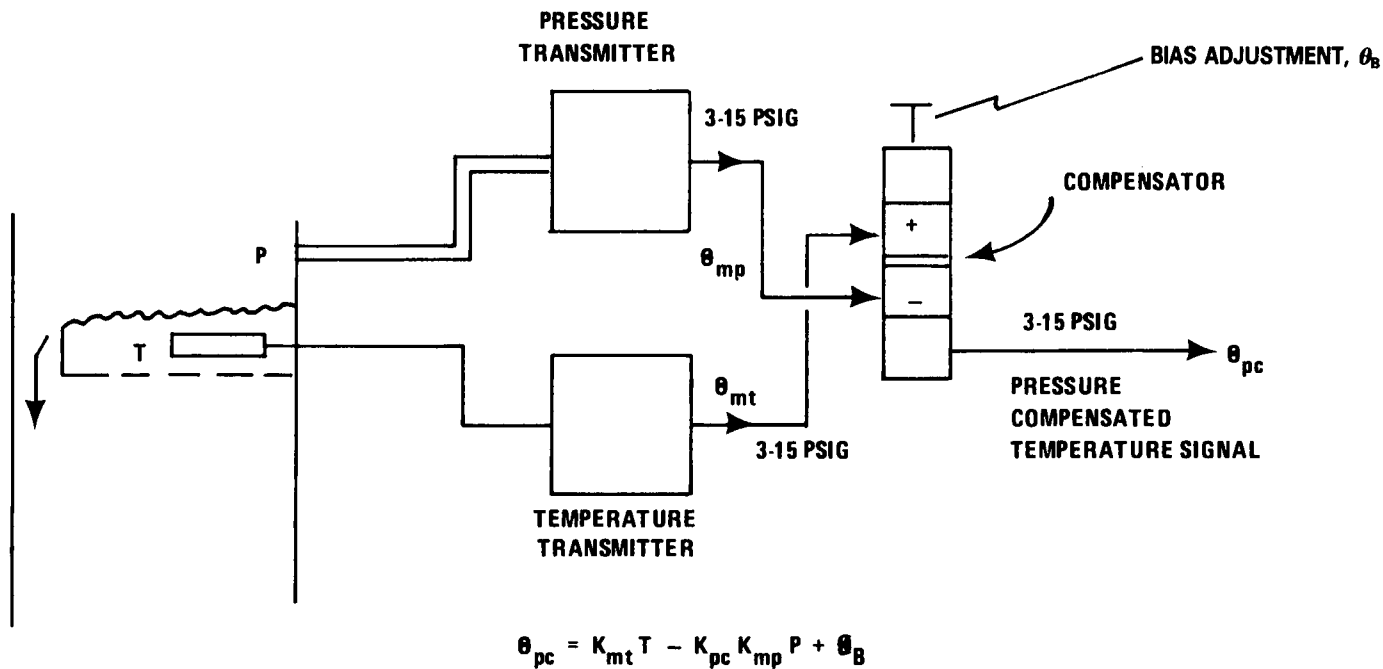


FIGURE 10.2
Pressure compensation of temperature measurement

The compensator has the equation:

$$\theta_{pc} = \theta_{mt} - K_{pc} \theta_{mp} + \theta_b \quad (10.8)$$

where

θ_{pc} = compensator output signal

θ_{mp} = pressure transmitter output signal

θ_{mt} = temperature transmitter output signal

K_{pc} = compensator gain

θ_b = compensator bias

This function is readily performed by commercial proportioning relays or some proportional-only controllers.

Theory for Binary Distillation

With the above equation in mind, we can see that if there is a pressure change ΔP from P_{FS} that causes a temperature change ΔT , we want *no* change in θ_{pc} if composition has not changed. In other words:

$$\Delta \theta_{pc} = \Delta \theta_{mt} - K_{pc} \Delta \theta_{mp} = 0 \quad (10.9)$$

or

$$K_{pc} = \left. \frac{\Delta \theta_{mt}}{\Delta \theta_{mp}} \right|_{\text{constant composition}} \quad (10.10)$$

$$\Delta \theta_{mt} = K_{mt} \Delta T \quad (10.11)$$

$$\Delta \theta_{mp} = K_{mp} \Delta P \quad (10.12)$$

where

K_{mt} = temperature transmitter gain

$$= \frac{(\Delta \theta_{mt})_{\max}}{T_{\max} - T_{\min}} \quad (10.13)$$

$T_{\max} - T_{\min}$ = temperature transmitter input span

$(\Delta \theta_{mt})_{\max}$ = transmitter output span (whether pneumatic or electronic is immaterial)

K_{mp} = pressure transmitter gain

$$= \frac{(\Delta \theta_{mp})_{\max}}{P_{\max} - P_{\min}}$$

$P_{\max} - P_{\min}$ = transmitter input span

$(\Delta \theta_{mp})_{\max}$ = transmitter output span

Next let us look at the temperature–pressure–composition relationship of Figure 10.3. At desired flowsheet composition C_{FS} we have flowsheet pressure P_{FS} and flowsheet temperature T_{FS} . The slope of the curve at P_{FS} and T_{FS} is $\partial P/\partial T$. From this we can see that if at constant composition there is a pressure change ΔP that causes a temperature change ΔT , then:

$$\frac{\Delta T}{\Delta P} = \frac{1}{\partial P/\partial T} \quad (10.14)$$

or

$$\Delta T = \frac{\Delta P}{\partial P/\partial T} \quad (10.15)$$

Equation (10.15) can now be combined with equations (10.10), (10.11), and (10.12) to solve for required K_{pc} :

$$\begin{aligned} K_{pc} &= \left. \frac{\Delta \theta_{mt}}{\Delta \theta_{mp}} \right|_{\text{constant composition}} = \left. \frac{K_{mt} \Delta T}{K_{mp} \Delta P} \right|_{\text{constant composition}} \\ &= \frac{K_{mt} \left(\frac{\Delta P}{\partial P/\partial T} \right)}{K_{mp} \Delta P} \bigg|_{\text{constant composition}} = \frac{K_{mt}}{K_{mp}} \left(\frac{1}{\partial P/\partial T} \right) \end{aligned} \quad (10.16)$$

Theory for Multicomponent Distillation

This is the same as for the binary case except for calculating $\partial T/\partial P$.

The equilibrium pressure, P_T , at a specified temperature, T , for a constant composition multicomponent mixture is a function of the vapor pressure, mole

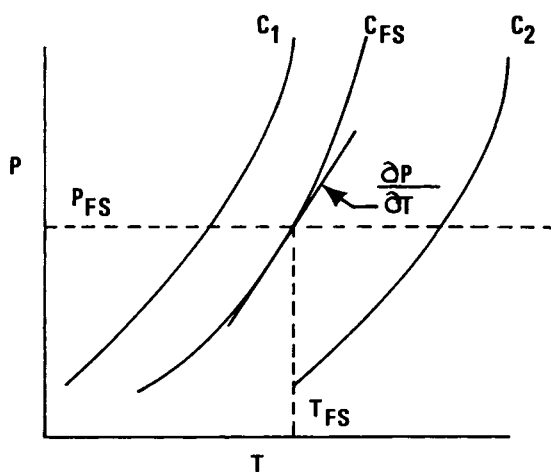


FIGURE 10.3
Composition vs. t and p

fraction, and activity coefficient for each component and is defined:

$$P_T = \sum_{i=1}^n \gamma_i x_i P_i$$

where n is the number of components and P_i is the vapor pressure at T for the i th component.

$$P_i = e^{[A_i + B_i/(T + 273.16)]}$$

so

$$P_T = \sum_{i=1}^n \gamma_i x_i e^{[A_i + B_i/(T + 273.16)]}$$

Differentiating, we get:

$$\begin{aligned} \frac{\partial P_T}{\partial T} &= -\frac{1}{(T + 273.16)^2} \sum_{i=1}^n \gamma_i x_i B_i e^{[A_i + B_i/(T + 273.16)]} \\ &= \frac{-1}{(T + 273.16)^2} \sum_{i=1}^n \gamma_i x_i B_i P_i \end{aligned}$$

and

$$\frac{\partial T}{\partial P_T} = \frac{1}{\partial P_T / \partial T}$$

Some Practical Considerations

1. The output of the compensator, θ_{pc} , is a measure of the pressure-compensated temperature. The process operator must know what this relationship is in order to adjust the set point for the controller properly. It is common practice to adjust the compensator bias so that it reads midscale when $T = T_{FS}$ and $P = P_{FS}$, or more generally when:

$$\theta_{mi} - K_{pc} \theta_{mp} = 0 \quad [\text{see equation (10.1)}]$$

Therefore:

$$\theta_B = \frac{(\Delta \theta_{pc})_{\max}}{2}$$

2. Some operators have found it confusing to have both temperature and pressure-compensated temperature. We therefore have taken Figure 10.3 and plotted C versus T (both read at constant P_{FS}) on a new plot. The operator's control station can then be calibrated directly in terms of composition.

3. In some cases the transmitted pressure signal is so noisy that a snubber may be required. For electronics an RC filter may be used.

If a computer is available, it is possible to take a considerably more accurate approach.

For a binary system, the basic equation is:

$$P_T = x_L P_L \gamma_L + (1 - x_L) P_H \gamma_H \quad (10.17)$$

where

γ_L = activity coefficient of low boiler

γ_H = activity coefficient, high boiler

Equation (10.17) may be solved for x_L :

$$x_L = \frac{P_T - P_H \gamma_H}{P_L \gamma_L - P_H \gamma_H} \quad (10.18)$$

The individual vapor pressures may be found by the Antoine equations; see equations (10.6) and (10.7). The activity coefficients may be found by equations (10.4) and (10.5).

In looking at equations (10.18), (10.4), and (10.5), we can see that x_L is really an implicit function. By knowing T we can readily calculate P_L and P_H , but from there on we must find x_L by trial and error. This is easily done on any computer.

10.6 MULTICOMPONENT COMPOSITIONS COMPUTED FROM TEMPERATURE AND PRESSURE MEASUREMENTS

If an on-line digital computer is available, it is possible to use a more sophisticated computing scheme than otherwise would be practical. To treat this subject in a very general way is not feasible. To show what is possible, however, let us consider as an example a three-component system where the ratio of the concentration of two components is essentially constant:

$$\frac{x_1}{x_2} = R \quad (10.19)$$

$$P = \sum_{i=1}^3 x_i \gamma_i P_i \quad (10.20)$$

$$P_i = \exp\left(A_i + \frac{B_i}{T + 273.16}\right) \quad (10.21)$$

It is desired to find x_1 from measurements of P_T and T , taken at the same tray in the column. We may rewrite equation (10.20):

$$x_1 = \frac{P - \gamma_3 P_3}{\gamma_1 P_1 + \frac{\gamma_2}{R} P_2 - \left(1 + \frac{1}{R}\right) \gamma_3 P_3} \quad (10.22)$$

Knowing A_i , B_i , one may find P_i from equation (10.21). Then knowing P_i and γ_i , one may find x_1 from equation (10.22). The activity coefficients

(γ_i 's) are sometimes fairly constant (if the range of composition changes is small). More generally, the γ 's would be functions of composition, so an iterative, trial-and-error solution would be required.

10.7 DOUBLE-DIFFERENTIAL TEMPERATURE

Strictly speaking one temperature and one pressure at a point in a column can be used to predict composition accurately only if the system is binary. The thought has occurred to a number of people that perhaps additional temperature measurements could be used to calculate composition in multicomponent systems. Two papers that illustrate two different approaches have appeared in the literature.

Luyben⁸ made a computer study of a 25-tray deisobutanizer with temperature measurements on trays 5, 10, and 15 (numbering up from the bottom). He defined:

$$\begin{aligned}\Delta^2 T &= (T_{10} - T_{15}) - (T_5 - T_{10}) \\ &= 2 T_{10} - T_5 - T_{15}\end{aligned}\quad (10.23)$$

Plots of $\Delta^2 T$ versus x_B (mol fraction iC_4) showed that it was affected very little by disturbances, but they did show a maximum at about $x_B = 0.02$. Since most deisobutanizers operate in a range of $x_B = 0.03 - 0.10$, this might well be useful for control.

Boyd⁵ took a somewhat different approach. He located two temperature measurements in the rectifying section of the column and two more in the stripping section. He then defined:

$$\Delta^2 T = \Delta T_R - \Delta T_S \quad (10.24)$$

where

ΔT_R = temperature difference in rectification section

ΔT_S = temperature difference in stripping section

Plant tests on two columns showed great sensitivity. In each case $\Delta^2 T$ was used to control top product; each system was calibrated by gas chromatographs. The computed and measured data checked well.

Boyd's technique partially fixed the composition-versus-tray profile in the vicinity of the feed tray and on both sides of it. Luyben's method partially fixed the profile in either end of the column. Both approaches effectively cancel out the influence of boilup rate and pressure on temperature measurements. Therefore, they both may be regarded as improvements on simple ΔT measurements, and are primarily useful for binary or almost-binary systems.

Bonilla⁶ reported that Boyd's double-differential-temperature technique did not work well for a C_3/C_4 splitter. The signal was found to be dependent on feed composition, and also was nonmonotonic.

10.8 AVERAGE TEMPERATURE

The use of multiple temperature measurements was proposed by Grote⁹ and independently (but much later) by Luyben.¹⁰ This technique is useful for distillation columns in which a large temperature change occurs over a few trays (a sharp temperature profile). This is found in columns where the components have widely differing boiling points (peanut butter and hydrogen would be an extreme example).

Control problems are frequently encountered in these columns because the process gain is very high (a small change in heat input makes a drastic change in the temperature on any single control tray). In addition, the system saturates easily: the temperature drops to the boiling point of the low boiling component as the profile drops below the control point but shows no change thereafter as the light component works down the column.

The multiple-temperature-control scheme uses a number (four or five) of temperature sensors located around the normal temperature break. An average temperature signal is computed from these multiple temperatures (this can be inexpensively implemented in conventional pneumatic or electronic analog hardware).

As the profile moves up or down, this average temperature changes gradually. The average temperature signal is fed into a conventional feedback controller. This technique has been successfully applied to many industrial columns.

10.9 COMPOSITION ESTIMATORS

Brosilow and co-workers¹¹⁻¹³ studied the use of several temperature and flow-rate measurements to estimate product composition. They used the term "inferential control" for this type of composition estimation and control using secondary process measurements.

The Brosilow estimator employs a linear combination of selected tray temperatures and steam and reflux flow rates to estimate product compositions. For example, distillate composition is estimated using an equation of the form:

$$x_{Dj} = a_{j1}T_1 + a_{j2}T_2 + a_{j3}T_3 + a_{j4}F_S + a_{j5}L_o \quad (10.25)$$

where

F_S = steam flow rate

L_o = external reflux flow rate

The a coefficients are constants that are determined experimentally or by calculations.

The estimator uses steady-state relationships, handles multicomponent systems, and is linear. Both simulation studies and experience on an industrial column have been reported. Shah¹⁴ had good success with the Brosilow estimator as long as the column was operated within a linear region. However, when nonlinear

phenomena occur, the linear estimator performs poorly; that is, poor estimates of product compositions are made.

Shah¹⁴ extended Brosilow's work, proposing a nonlinear composition estimator for binary systems. A fundamental nonlinear mathematical model of the column is used, involving tray-to-tray material and energy balances, vapor-liquid equilibrium and physical property data, known process parameters (tray efficiencies and heat losses), and measured variables (feed, distillate, and steam flow rates and one or more tray temperatures).

Suppose, for example, that reflux and distillate flow rates and a tray temperature in the rectifying section are known, and one wants to calculate distillate composition x_D . The rigorous estimator makes a guess of x_D and calculates tray-by-tray down the column until the tray is reached on which the temperature is measured. If the calculated temperature on this tray does not agree with the measured temperature, a new guess of x_D is made and the procedure is repeated.

Shah verified the effectiveness of the nonlinear estimator by both simulation studies and experimental tests on a pilot-scale column. The work of Weber and Mosler¹⁵ should also be noted. They proposed several control schemes using the sum of two or more temperatures, the difference between two temperatures, and the distillate-to-feed ratio to adjust manipulated variables.

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11

Miscellaneous Measurements and Controls

11.1 INTRODUCTION

We make no effort to cover all measurements and hardware here. We have selected, however, a number of items we think are important, or are worth emphasizing, or have not been covered adequately elsewhere. A glaring omission is that of flow metering. We are very suspicious of the accuracy of most in-plant flow meters because only rarely do plants have facilities for their calibration, but we have nothing new or unique to offer for improvement. Considerable emphasis, on the other hand, has been placed on temperature-measurement dynamics since this subject seems to be widely misunderstood or inadequately understood. And we have devoted some space to control-valve sizing and selection since the new (1974) ISA equations for calculating the operating C_v and for predicting the flow regime are not yet widely used.

11.2 CALCULATION OF DISTILLATION-COLUMN INTERNAL REFLUX

In the design of distillation columns, engineers use internal reflux in their calculations rather than external reflux. Important indices for column separation ability are internal reflux/distillate and boilup/bottom-product ratios.

Increasingly we are controlling columns on the basis of internal reflux rather than external reflux. The most common method of determining internal reflux makes use of external reflux flow measurement and the number of degrees of condensate subcooling* (see Figure 11.1 and reference 15):

* This method works fairly well for relatively pure material. For mixtures of components, it is sometimes advisable to take into account the difference between bubble point and dew point.

$$w_{R1} = w_R \left[1 + \frac{c_p}{\lambda} (T_O - T_R) \right] \quad (11.1)$$

$$= w_R K_{SC} \quad (11.1a)$$

where

w_{R1} = internal reflux rate, lbm/hr

w_R = external reflux rate, lbm/hr

c_p = liquid specific heat, pcu/lbm °C

λ = latent heat of vaporization of vapor
in column, pcu/lbm

T_O = vapor temperature, °C

T_R = condensate (external reflux) temperature,
°C at point of entry to the column

To implement equation (11.1), we need a summer with adjustable gain and bias, and a multiplier. For the former the Foxboro 136-1 is a suitable device† while for the latter we would use the Foxboro 556-8 multiplier. The discussion that follows, however, could be readily extended to other pneumatic, electronic, or digital devices. A commonly used hardware arrangement is shown in Figure 11.2.

Summer

The Foxboro 136-1 summer has the equation:

$$p = K_R (A - C) + B \quad (11.2)$$

where

p = output pressure, psig

A = signal from vapor temperature
transmitter, psig

C = signal from external reflux (condensate)
temperature transmitter, psig

B = bias, psig

K_R = summer gain, psi/psi

It is assumed, incidentally, that the two temperature transmitters have the same span, and therefore the same gain:

$$K_{mt} = \frac{12 \text{ psi}}{(\Delta T)_{\text{span}}} \quad (11.3)$$

† There are several satisfactory pneumatic summers on the market, but those that use pressure-dividing networks for gain should be avoided. They are not accurate enough.

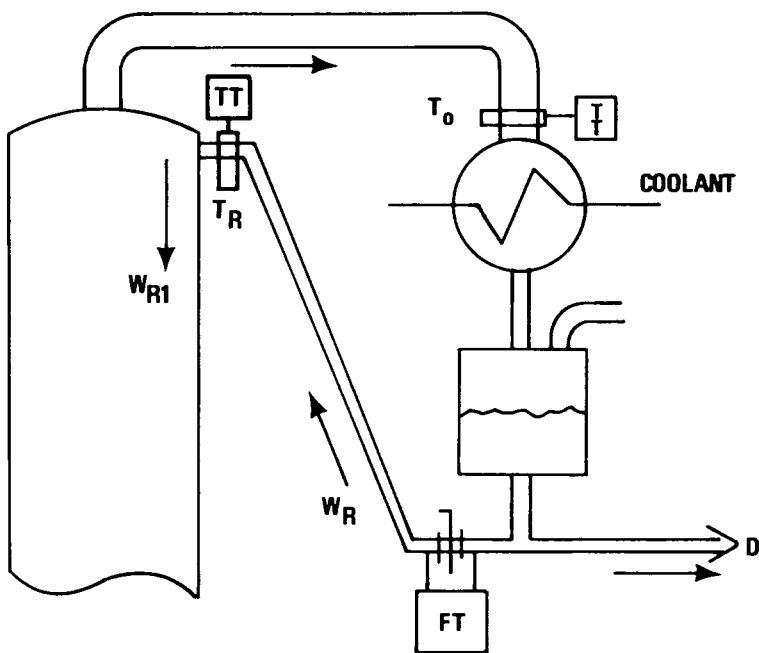


FIGURE 11.1
Measurements needed for internal reflux computation

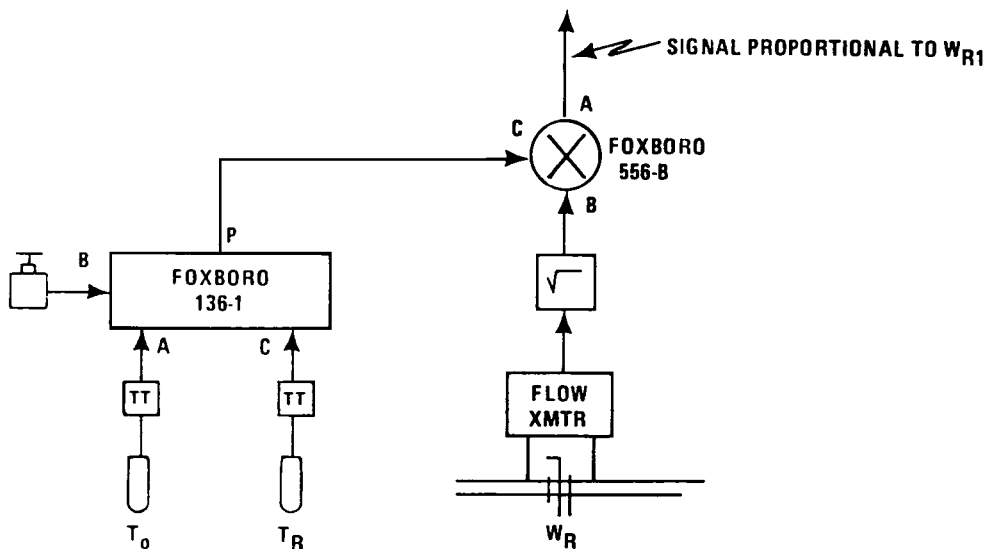


FIGURE 11.2
Pneumatic hardware configuration for internal reflux computation

Next we must have a procedure for determining B and K_R . Let us assume that when $T_O - T_R$ is some specified maximum value, $(T_O - T_R)_{\max}$, we want $p = 15$ psig. Let us also say that we want the output span of the summer to be 12 psig for 0 to $[K_{SC}]_{\max}$. When the two temperatures are equal—no subcooling— K_{SC} will equal $1.00 = [K_{SC}]_{\min}$. A will equal C , so equation (11.2) becomes:

$$p = B \quad (11.4)$$

Since $0 - [K_{SC}]_{\max}$ corresponds to 12 psi, and B corresponds to $[K_{SC}]_{\min} = 1$, then:

$$B = \frac{[K_{SC}]_{\min}}{[K_{SC}]_{\max}} \times 12 + 3 = \frac{1}{[K_{SC}]_{\max}} \times 12 + 3 \quad (11.5)$$

When $T_O - T_R$ is maximum, $A - C$ is maximum and

$$K_R (A - C)_{\max} = \frac{[K_{SC}]_{\max} - 1}{[K_{SC}]_{\max}} \times 12 \quad (11.6)$$

Also:

$$(A - C)_{\max} = (T_O - T_R)_{\max} K_{mt} \quad (11.7)$$

Thus by combining equations (11.6) and (11.7) and solving for K_R , we get:

$$K_R = \frac{[K_{SC}]_{\max} - 1}{[K_{SC}]_{\max}} \times \frac{12}{(T_O - T_R)_{\max} \times K_{mt}} \quad (11.8)$$

$$= \frac{(c_p/\lambda)(T_O - T_R)_{\max}}{1 + \frac{c_p}{\lambda}(T_O - T_R)_{\max}} \times \frac{12}{(T_O - T_R)_{\max} K_{mt}} \quad (11.8a)$$

$$= \frac{12 (c_p/\lambda)}{\left[1 + \frac{c_p}{\lambda}(T_O - T_R)_{\max} \right] K_{mt}} = \frac{(c_p/\lambda)(\Delta T)_{\text{span}}}{[K_{SC}]_{\max}} \quad (11.8b)$$

Multiplier

The Foxboro 556-8 multiplier has the equation:

$$A - 3 = \frac{f}{12} (B - 3) [(1 - S)12 + S(C - 3)] \quad (11.9)$$

Let us put the signal from the external reflux flow transmitter into the B connection and the signal from the summer into the C connection. Let us further assume that the output of the multiplier, proportional to the internal reflux, has the same span as that of the external reflux flow transmitter (assumed linear). We now need to find the f and S factors.

When the external reflux is:

$$\frac{(w_R)_{\max}}{2}$$

$$B - 3 = 6.00 \text{ psi}$$

If at the same time the temperature difference is maximum, we want:

$$A - 3 = 6.00 \times [K_{SC}]_{\max} \quad [\text{from equation (11.1a)}] \quad (11.10)$$

$$= \frac{f}{12} \times 6.00 [(1 - S) \times 12 + 12S] \quad (11.11)$$

$$= 6.00 f \quad (11.11a)$$

Note that $C - 3 = 12$ since:

$$K_{SC} = [K_{SC}]_{\max}$$

Next if external reflux is:

$$\frac{(w_R)_{\max}}{2}$$

and if the temperature difference is zero:

$$A - 3 = 6.00 [K_{SC}]_{\min} = 6.00 \quad (11.12)$$

$$= \frac{f}{12} \times 6.00 \left[(1 - S) \times 12 + \frac{12S}{[K_{SC}]_{\max}} \right] \quad (11.13)$$

Note that for $A - C = 0$:

$$p = B = \frac{12}{[K_{SC}]_{\max}} + 3 = C$$

Equations (11.10) and (11.11a) may now be solved to find f :

$$f = \frac{6.00 [K_{SC}]_{\max}}{6.00} = [K_{SC}]_{\max} \quad (11.14)$$

This value of f may be substituted into equations (11.12) and (11.13) to find S :

$$6.00 = [K_{SC}]_{\max} \times 0.5 \left[(1 - S) \times 12 + \frac{12}{[K_{SC}]_{\max}} S \right] \quad (11.15)$$

$$= (1 - S) 6.00 [K_{SC}]_{\max} + 6.00 S \quad (11.16)$$

$$1 = [K_{SC}]_{\max} - [K_{SC}]_{\max} S + S \quad (11.17)$$

whence

$$S = 1 \quad (11.18)$$

Example. Let us consider a case where the following conditions apply:

External reflux flow-meter span = 0–50,000 pph

Temperature transmitter spans = 0–150°C

$c_p = 0.71$ pcu/lbm °C; $(T_O - T_R)_{\max} = 100^\circ\text{C}$

$\lambda = 235$ pcu/lbm

From these:

$$\begin{aligned}[K_{SC}]_{\max} &= 1 + \frac{0.71}{235} \times 100 \\ &= 1.302 = f\end{aligned}$$

From equation (11.5):

$$\begin{aligned}B &= \frac{12}{1.302} + 3.00 \\ &= 9.22 + 3.00 \\ &= 12.22 \text{ psig}\end{aligned}$$

From equation (11.8b):

$$\begin{aligned}K_R &= \frac{12 \times \frac{0.71}{235}}{1.302 \frac{12}{150}} \\ &= 0.348\end{aligned}$$

From equation (11.14), $f = 1.302$, and from equation (11.18), $S = 1$.

As a check let:

$$w_R = 25,000 \text{ pph}$$

$$T_O - T_R = 40^\circ\text{C}$$

$$\begin{aligned}p &= 0.348 \times \frac{40 \times 12}{150} + 12.22 \quad [\text{from equation (11.2)}] \\ &= 1.1136 + 12.22 = 13.33 \text{ psig} \\ &= C\end{aligned}$$

Then:

$$\begin{aligned}A - 3 &= \frac{1.302}{12} \times 6.00 \quad (10.33) \\ &= 6.725\end{aligned}$$

Therefore:

$$w_{R1} = \frac{6.725}{12} \times 50,000 = 28,022 \text{ pph}$$

Check via equation (11.1):

$$\begin{aligned} w_{R1} &= 25,000 \left(1 + \frac{0.71}{235} \times 40 \right) \\ &= 28,021 \text{ pph} \end{aligned}$$

11.3 TEMPERATURE AND PRESSURE COMPENSATION OF GAS FLOW METERS

The basic gas-flow metering equation is:

$$q = k_1 \sqrt{\frac{\Delta h_o (p + 14.7)}{(T + 273)(\rho_g/\rho_a)SC}} \quad (11.19)$$

where

Δh_o = orifice pressure drop, inches w.c.

p = pressure, psig. It should be measured upstream if orifice reference pressure p_c is taken upstream, and downstream if p_c is taken downstream.

T = temperature, °C, of flowing stream where p is measured

ρ_g/ρ_a = ratio of gas density to that of air at 14.7 psia, 60°C

SC = gas supercompressibility

k_1 = orifice flow constant

q = flow, scfm or pph

Let us also define:

p_c = orifice reference pressure, psia, used in designing orifice

T_c = orifice reference temperature, °C, used in designing orifice

p_s = pressure transmitter span, psi

T_s = temperature transmitter span, °C

For a particular value of Δh_o measured at T_c and p_c :

$$q_c = k_1 \sqrt{\frac{\Delta h_o (p_c + 14.7)}{(T_c + 273)(\rho_g/\rho_a)SC}} \quad (11.20)$$

$$= k_2 \sqrt{\Delta h_o} \quad (11.21)$$

For the same Δh_o but $p \neq p_c$ and $T \neq T_c$, we can divide equation (11.20) to get:

$$\frac{q}{q_c} = \sqrt{\frac{(T_c + 273)(p + 14.7)}{(T + 273)(p_c + 14.7)}} \quad (11.22)$$

Substituting equation (11.21) into equation (11.22), we obtain:

$$q = k_2 \sqrt{\frac{\Delta h_o (T_c + 273)(p + 14.7)}{(T + 273)(p_c + 14.7)}} \quad (11.23)$$

As can be seen, equation (11.23) provides a convenient way of determining true flow when $p \neq p_c$ and $T \neq T_c$.

A common method of mechanizing equation (11.23) with pneumatic instruments is given in Figure 11.3 (see Foxboro *TI 45-3a* of October 14, 1966). A more recent Foxboro bulletin, *TI 45-3a* of November 1969, features two alternate arrangements. We have been unable to find in them any real advantage over the arrangement of Figure 11.3, so the following will apply to Figure 11.3.

Temperature Compensation

The usual divider form of the 556 is:

$$B - 3 = \frac{12}{f} \times \frac{(A - 3)}{12Z + S(C - 3)} \quad (11.24)$$

and Foxboro defines for this application:

$$f = \frac{(T_m)_{\max} + 273}{T_c + 273} \quad (11.25)$$

$$S = \frac{(T_m)_{\max} - (T_m)_{\min}}{(T_m)_{\max} + 273} \quad (11.26)$$

$$= \Delta T_s / [(T_m)_{\max} + 273] \quad (11.27)$$

$$Z = 1 - S$$

where

$(T_m)_{\max}$ = maximum measured temperature, °C

$(T_m)_{\min}$ = minimum measured temperature, °C

$$(T_m)_{\max} - (T_m)_{\min} = \Delta T_s$$

= temperature transmitter
span, °C

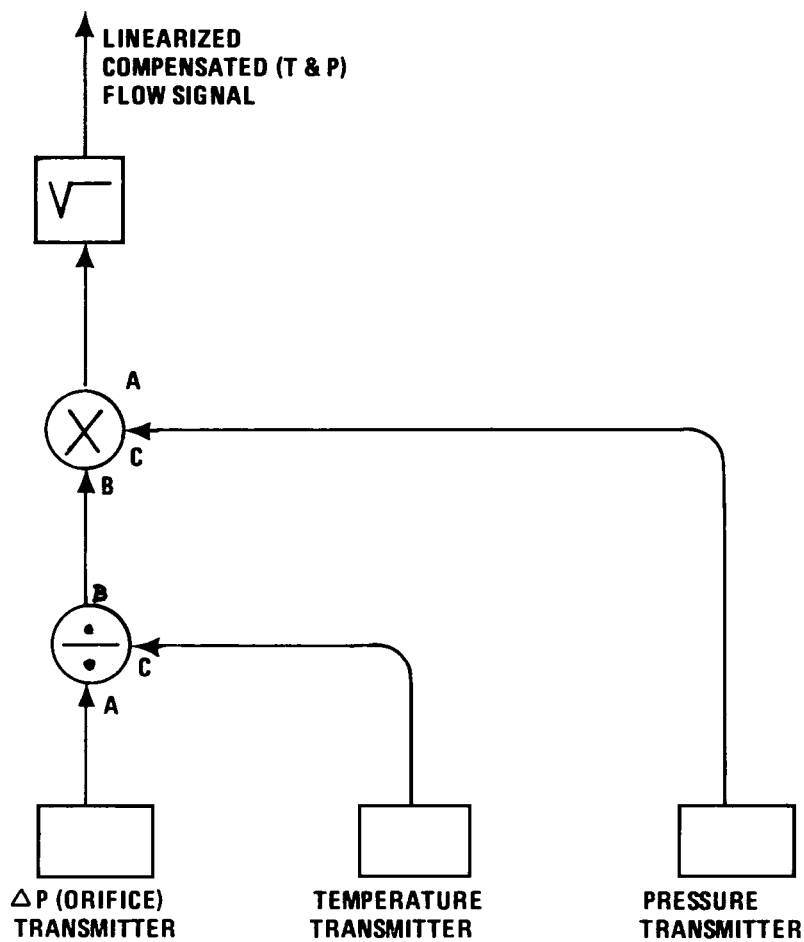


FIGURE 11.3
Typical compensated gas flow metering scheme

A is the port receiving the Δh_o transmitter signal, θ_{mf} , and C is the port receiving the temperature transmitter signal, θ_{mt} . On substituting equations (11.25), (11.26), and (11.27) into equation (11.24), we get:

$$B - 3 = \frac{(T_c + 273) \times 12}{(T_m)_{\max} + 273} \times \frac{\theta_{mf} - 3}{\left[12 \left[1 - \frac{\Delta T_s}{(T_m)_{\max} + 273} \right] + \frac{\Delta T_s}{(T_m)_{\max} + 273} \times (\theta_{mt} - 3) \right]} \quad (11.28)$$

$$= \frac{(T_c + 273)(\theta_{mf} - 3)}{[(T_m)_{\min} + 273] + \Delta T_s \left(\frac{\theta_{mf} - 3}{12} \right)} \quad (11.29)$$

where

θ_{mf} = signal from Δh_o transmitter

θ_{mt} = signal from temperature transmitter

It can be seen that the denominator of equation (11.29) is equal to $T + 273$. This calculation, therefore, takes care of span suppression and converts both T and T_c to absolute units as required.

Some engineers omit temperature compensation of steam flow on the grounds that it changes more slowly than pressure. The wisdom of this is questionable.

Pressure Compensation

The multiplier form of the 556 is:

$$A - 3 = \frac{f}{12} \times (B - 3) \times [12Z + S(C - 3)] \quad (11.30)$$

and for this application Foxboro defines:

$$f = \frac{(p_m)_{\max} + 14.7}{p_c + 14.7} \quad (11.31)$$

$$S = \frac{(p_m)_{\max} - (p_m)_{\min}}{(p_m)_{\max} + 14.7} \quad (11.32)$$

$$Z = 1 - S \quad (11.33)$$

where

$(p_m)_{\max}$ = maximum value of transmitter span, psig

$(p_m)_{\min}$ = minimum value of transmitter span, psig

$$(p_m)_{\max} - (p_m)_{\min} = \Delta p_s$$

= pressure transmitter span, psi

B is the port where the signal from the divider enters, and C is the port where the pressure transmitter signal θ_{mp} enters.

Substituting equation (11.31), (11.32), and (11.27) into equation (11.30):

$$A - 3 = \frac{(p_m)_{\max} + 14.7}{12 [p_c + 14.7]} \times (B - 3) \left[12 \left[1 - \frac{\Delta p_s}{(p_m)_{\max} + 14.7} \right] + \frac{\Delta p_s}{(p_m)_{\max} + 14.7} (\theta_{mp} - 3) \right] \quad (11.34)$$

where

θ_{mp} = signal from pressure transmitter

Clearing, we obtain:

$$A - 3 = \frac{(B - 3)}{p_c + 14.7} \left[[(p_m)_{\min} + 14.7] + \Delta p_s \frac{(\theta_{mp} - 3)}{12} \right] \quad (11.35)$$

But:

$$[(p_m)_{\min} + 14.7] + \Delta p_s \frac{(\theta_{mp} - 3)}{12} = p + 14.7 \quad (11.36)$$

so:

$$A - 3 = \left(\frac{p + 14.7}{p_c + 14.7} \right) (B - 3) \quad (11.37)$$

This procedure, therefore, takes care of span suppression and converts both p and p_c to absolute units.

Substituting for $B - 3$ from equation (11.29), we get:

$$A - 3 = (\theta_{mf} - 3) \frac{(T_c + 273)(p + 14.7)}{(T + 273)(p_c + 14.7)} \quad (11.38)$$

Since $A - 3$ is now a measure of q^2 , we follow the multiplier with a square root extractor to get a signal proportional to flow that is corrected for pressure deviations from p_c and for temperature deviations from T_c .

In specifying Foxboro 556-8's and 556-9's, one must be careful to stay within available ranges for f and span S . This sometimes requires juggling transmitter spans.

Errors Incurred by Uncompensated Gas Flows

For estimates of flow-metering errors that may occur when compensation is not made for temperature and pressure deviations from reference values, we

may take the partial derivative of equation (11.19) with respect to temperature and pressure.

$$(\Delta q)_T = \frac{\partial q}{\partial T} \Delta T \quad (11.39)$$

$$\frac{\partial q}{\partial T} = k_1 \sqrt{\frac{\Delta h_o (p + 14.7)}{(\rho_g/\rho_a)SC}} \times \left(-\frac{1}{2}\right) \times \frac{1}{(\bar{T} + 273)^{3/2}} \quad (11.40)$$

$$= \frac{-\bar{q}}{2(\bar{T} + 273)} \quad (11.41)$$

$$\text{Percent error} = \frac{-100}{2(\bar{T} + 273)} \times \Delta T \quad (11.42)$$

Next:

$$(\Delta q)_p = \frac{\partial q}{\partial p} \Delta p \quad (11.43)$$

$$\frac{\partial q}{\partial p} = k_1 \sqrt{\frac{\Delta h_o}{(T + 273)(\rho_g/\rho_a)SC}} \times \frac{1}{2} \times \frac{1}{\bar{p} + 14.7} \quad (11.44)$$

$$= \frac{-\bar{q}}{2(\bar{p} + 14.7)} \quad (11.45)$$

$$\text{Percent error} = \frac{100}{2(\bar{p} + 14.7)} \times \Delta p \quad (11.46)$$

Consider, for example, the following:

Steam flow = 1500 pph

Design temperature = 397°F

Design pressure = 240 psia

Δh_o is assumed to be constant.

Example 1

Operating temperature = 397°F

Operating pressure = 215 psia

$$\frac{1500}{2 \times 240}(-25) = -78 \text{ pph}$$

= error in steam
flow-meter reading

$$\frac{-78 \times 100}{1500} = -5.2 \text{ percent error}$$

Example 2

Operating temperature = 437°F

Operating pressure = 240 psia

$$\frac{-1500}{2 \times (397 + 460)} \times 40 = -35 \text{ pph}$$

= error in steam
flow-meter reading

$$\frac{-35 \times 100}{1500} = -2.3 \text{ percent error}$$

11.4 HEAT-FLOW COMPUTATIONS

As mentioned in Chapters 3 and 4, it is sometimes desirable to calculate heat flow, reflux rate, or boilup rate. Let us consider, for example, the heat exchanger of Figure 11.4. We wish to know the increase or decrease of enthalpy

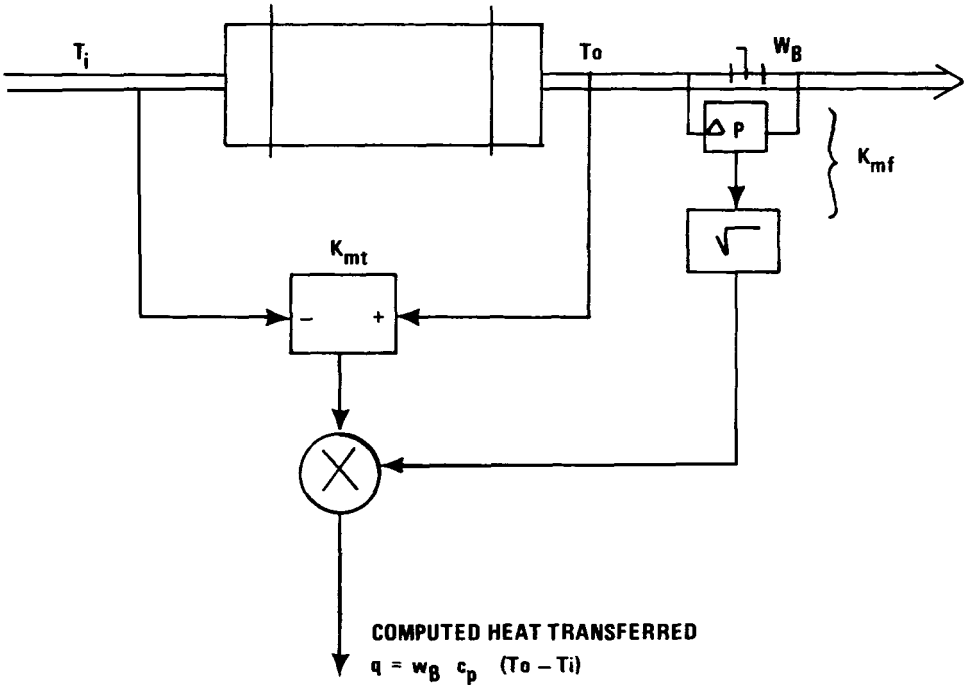


FIGURE 11.4
Heat flow computer for heat transfer

of stream B . If we know the flow rate w_B and the two temperatures T_i and T_o , we may calculate the rate of heat transfer:

$$q = w_B c_p (T_o - T_i) \quad (11.47)$$

Let:

$$q_{\text{meas}} = \text{heat-transfer span, pcu/hr}$$

$$(w_B)_{\text{max}} = \text{maximum flow-meter span, lbm/hr}$$

$$(T_o - T_i)_{\text{span}} = \Delta T_{\text{span}}, ^\circ\text{C}$$

For the Foxboro 556-8, it readily may be shown that:

$$f = \frac{(w_B)_{\text{max}} c_p (T_o - T_i)_{\text{span}}}{q_{\text{meas}}} \quad (11.48)$$

In many cases we are not as interested in heat transferred per hour. For a condenser we would probably like to know w_c , lbm/hr condensed. Then:

$$f = \frac{(w_B)_{\text{max}} c_p (T_o - T_i)_{\text{span}}}{(w_c)_{\text{max}} \lambda} \quad (11.49)$$

where

$$(w_c)_{\text{max}} = \text{span for estimated lbm/hr condensed}$$

$$\lambda = \text{latent heat of condensing vapor, pcu/lbm}$$

In this case w_B is the coolant flow and $(T_o - T_i)$ is its temperature rise.

Similarly, for reboilers heated with hot oil, we will want to know and control the rate of boilup, w_{BU} :

$$f = \frac{(w_B)_{\text{max}} c_p (T_i - T_o)_{\text{span}}}{(w_{BU})_{\text{max}}} \quad (11.50)$$

In this case w_B is the hot-oil flow and $(T_i - T_o)$ is the temperature drop.

A common method of measuring $T_o - T_i$ is to connect two resistance thermometers differentially. The output of the converter or relay is then a measure of $T_o - T_i$.

For the above applications, it is immaterial whether one puts the flow signal into the B port and the ΔT signal into the C port of the Foxboro 556-8, or vice versa.

11.5 COLUMN-BASE LEVEL MEASUREMENT¹

Introduction

Many problems have been encountered in making head measurements in and around distillation columns, that is, column ΔP , liquid level, and specific gravity. Of these column-base level is probably the one where difficulties are

most often encountered. It is the subject of this section. These difficulties are at least partially process oriented in the sense that a column base with associated reboiler is often very complex. Heat must be transferred, liquid must be partially vaporized, and liquid and vapor must be properly separated. In the discussion that follows, we will assume that process design is adequate. We will be concerned only with problems associated with determining level by means of head measurements.

Regardless of the type of transmitter used— ΔP or displacer—the basic equation relating liquid level, density, and head is (see Figure 11.5):

$$\Delta P_H = \frac{g_L}{g_c}(\rho_L - \rho_v)H + \frac{g_L}{g_c}\rho_v \Delta H_N \quad (11.51)$$

where

ΔP_H = head, lb force/ft², between two nozzles

g_L = local gravity acceleration, ft/sec²

g_c = mass-force conversion factor
= 32.2 ft lb mass/sec² lb force

ρ_L = liquid phase density, lb mass/ft³

ρ_v = vapor phase density, lb mass/ft³

H = liquid-vapor interface elevation above bottom nozzle, feet

ΔH_N = nozzle-to-nozzle spacing, feet

$$\Delta P_H = \frac{g_L}{g_c}(\rho_L - \rho_v)H + \frac{g_L}{g_c}\rho_v \Delta H_N$$

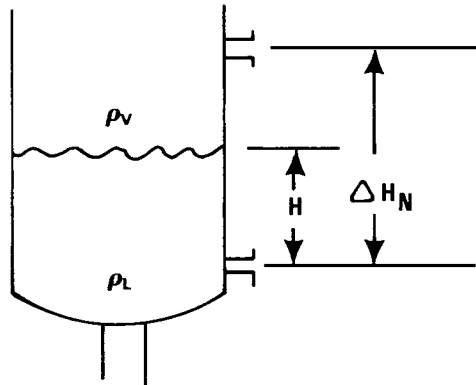


FIGURE 11.5
Head-level relationship in a vessel

The vapor density, ρ_v , is commonly assumed to be negligible, but this is not necessarily valid at pressures much above atmospheric. If, however, ρ_v is very small, equation (11.51) reduces to:

$$\Delta P_H = \frac{g_L}{g_c} \rho_L H \quad (11.52)$$

With the basic mathematics of level measurement in hand, let us next turn our attention to the problem of noise. Distillation-column base-level measurements tend to be extremely noisy. This is mostly attributable to turbulence, but may also be due to the type of transmitter used. Regardless of its source, however, it is best to filter noise out at the transmitter input rather than at its output. If this is done properly, such filtering or damping greatly reduces the probability of saturation in the transmitter and minimizes output signal errors due to transient, nonlinear transmitter operation. In the case of pneumatics, such filtering minimizes demands on the air supply and therefore minimizes the probability of "gulping."

Wave Noise in Large Columns

Predicting the amplitude and frequency of wave noise generally is very difficult. We can make a stab, however, at predicting the lowest frequency of standing waves by a method suggested in part by Binder.¹

This method says that the velocity of surface waves in a shallow environment is:

$$V_w = \sqrt{g_L \bar{H}} \quad (11.53)$$

where

V_w = surface wave velocity, ft/sec

g_L = gravitational constant, ft/sec²

\bar{H} = average liquid depth, feet

If we consider the column base to be a circular pool, then standing waves with the center as origin will exist at frequencies of:

$$f = 0, \frac{V_w}{2L}, \frac{V_w}{L}, \frac{3V_w}{2L}, \text{ etc.} \quad (11.54)$$

where

f = cycles per second (cps)

L = tower base radius, feet

As an example, consider a tower 17 feet in diameter with an average liquid depth of 5 feet. Then:

$$f = 0.75 \text{ cps, } 1.50 \text{ cps, } 2.25 \text{ cps, etc.}$$

Noise amplitude is usually greater at low frequencies. Therefore, in this case, any filtering or damping that we design should provide significant attenuation for frequencies as low as 0.75 cps.

Column-Base-Reboiler "Manometer"

Another source of noise is sometimes encountered in columns with thermosyphon reboilers. It has the natural frequency of the "manometer" between the column base and the reboiler (Figure 11.6). Roughly this is:

$$f_M = \frac{1.3}{\sqrt{L}} \text{ cps} \quad (11.55)$$

where

L = length of piping from base to reboiler, feet

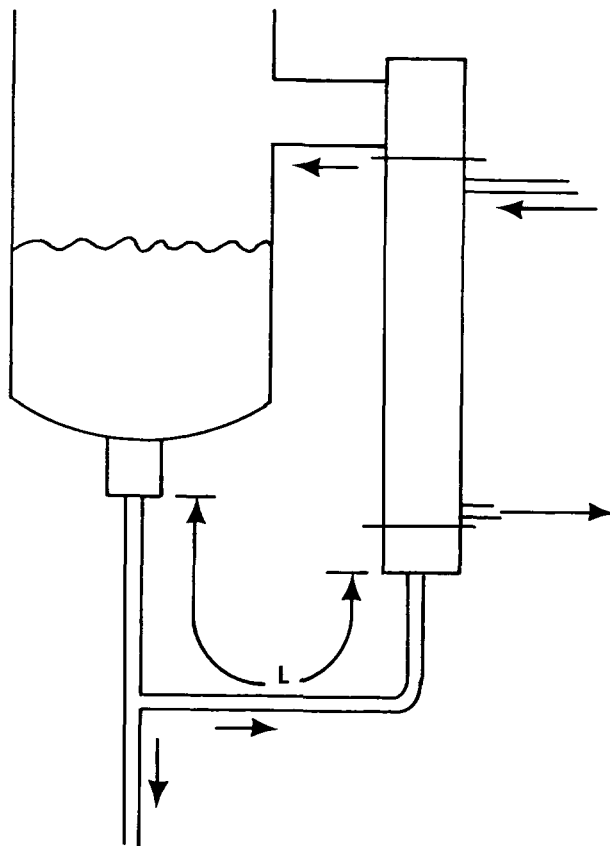


FIGURE 11.6
Column base-reboiler manometer

As an example, for 20 feet of piping:

$$f_M = \frac{1.3}{\sqrt{20}} = 0.3 \text{ cps}$$

The amplitude of this cycle may be arbitrarily attenuated by a properly sized restricting orifice in the line, or by snugly sized piping. The use of adjustable hand valves is highly undesirable; in many cases “fiddling” with such valves has seriously delayed column startup or has caused premature shutdowns.

Displacer-type Transmitter Resonance

A third source of noise—often the most serious—is attributable to the basic design of most displacer-type transmitters. The displacer and torque tube (Figure 11.7) constitute an open-loop, mass-spring system that is usually highly underdamped. As a consequence any transient disturbance causes the displacer to bounce up and down at a fairly well-determined frequency. The basic mathematics are given elsewhere (pages 187–191 of reference 2); for our immediate purposes it is sufficient to note that most commercial transmitters of this type have a natural frequency in the range of 1–3 cps.

There are now several force-balance, displacer-type transmitters on the market. The incomplete information available indicates that these have a higher natural frequency and more damping than do the nonforce balance type.

Internal Damping Chamber

As a first approach to noise filtering or damping, let us consider an internal damping chamber for use with either ΔP or displacer-type transmitters, which

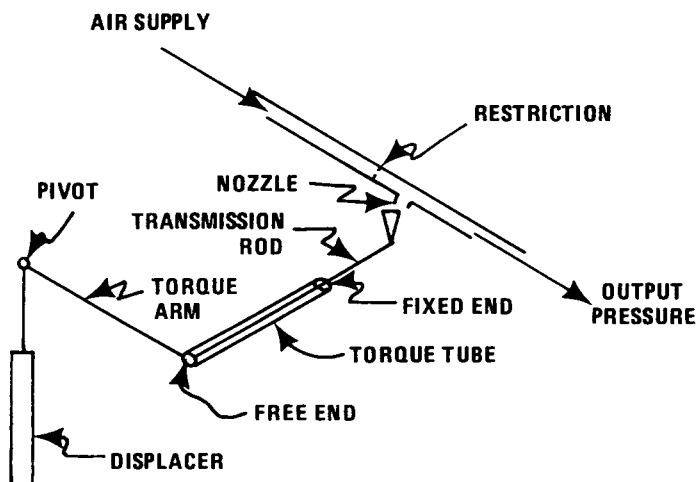


FIGURE 11.7
Schematic diagram of displacement-type level transmitter

functions as a first-order lag. As shown by Figure 11.8, this chamber has two key features: (1) a hood or cap to keep out liquid from the last downcomer or the reboiler return nozzle, and (2) a 1-inch-diameter hole in line with the lower level nozzle to facilitate rod out. The combination of orifice and chamber gives about 10:1 damping (attenuation) at about 0.5–0.6 cps. If there is a concern about solids collecting in the chamber, a second 1-inch hole may be cut in the lower section of the chamber. The damping will then be 10:1 at 1.0–1.2 cps.

It is probably apparent that almost any desired damping may be obtained by the proper choice of cross-sectional area in the chamber and of orifice-flow cross-sectional area.

External Damping for ΔP Transmitters

When a ΔP transmitter is used for level measurement and is installed at an elevation above the upper level tap and equipped with purges, one may provide damping as shown in Figure 11.9. Here a restriction and volume pot are installed in each impulse line to ensure that the inputs to the transmitter are dynamically equalized. Many combinations of volume pots and restrictors are possible; as an example, a pot with 12 in³ and a Taylor snubber (58S104) will give 10:1 attenuation at about 2 cps.

Useful for level measurements, this design is almost mandatory for specific gravity measurements via ΔP for slurries. It may also be used with gas or steam flow-meter installations with self-draining impulse lines.

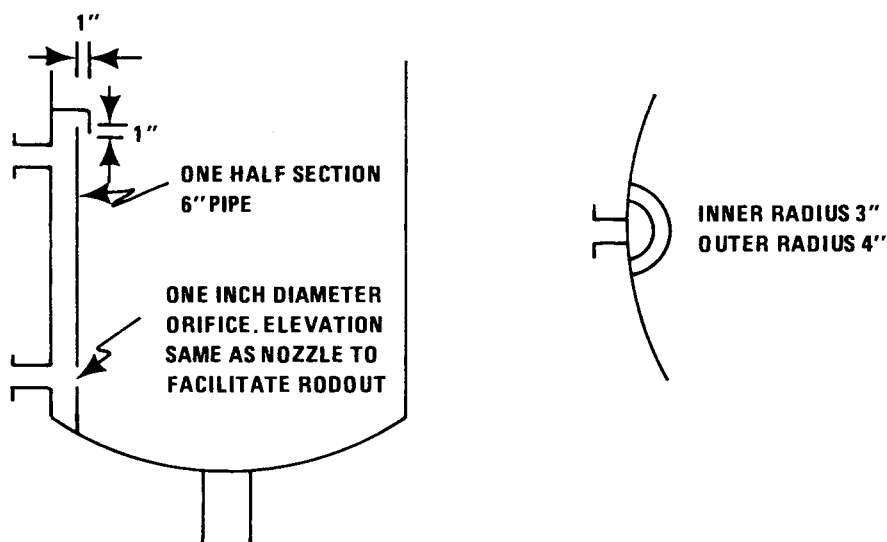


FIGURE 11.8
Internal damping chamber

The simple, first-order, filter approach may not always be adequate. In fact, if too large a time constant is chosen, it may interfere with control. A more sophisticated approach is to determine experimentally the noise-frequency spectrum and to design notch filters (i.e., band rejection filters) for the dominant frequencies.

ΔP Transmitters with High-Viscosity Fill

At least one commercial ΔP transmitter is available with a high-viscosity liquid fill. Its dynamics approximate those of a first-order lag with $\tau \approx 0.7$ second. The damping affects both input signals and the feedback, and thus minimizes pilot saturation and output errors. This instrument has the desirable feature of nonadjustable damping and so is "fiddleproof." Experience to date (10–15 years and several hundred applications) has been very favorable on flow-control applications, and a bit less favorable on level. Apparently many level applications, particularly in column bases, need additional damping.

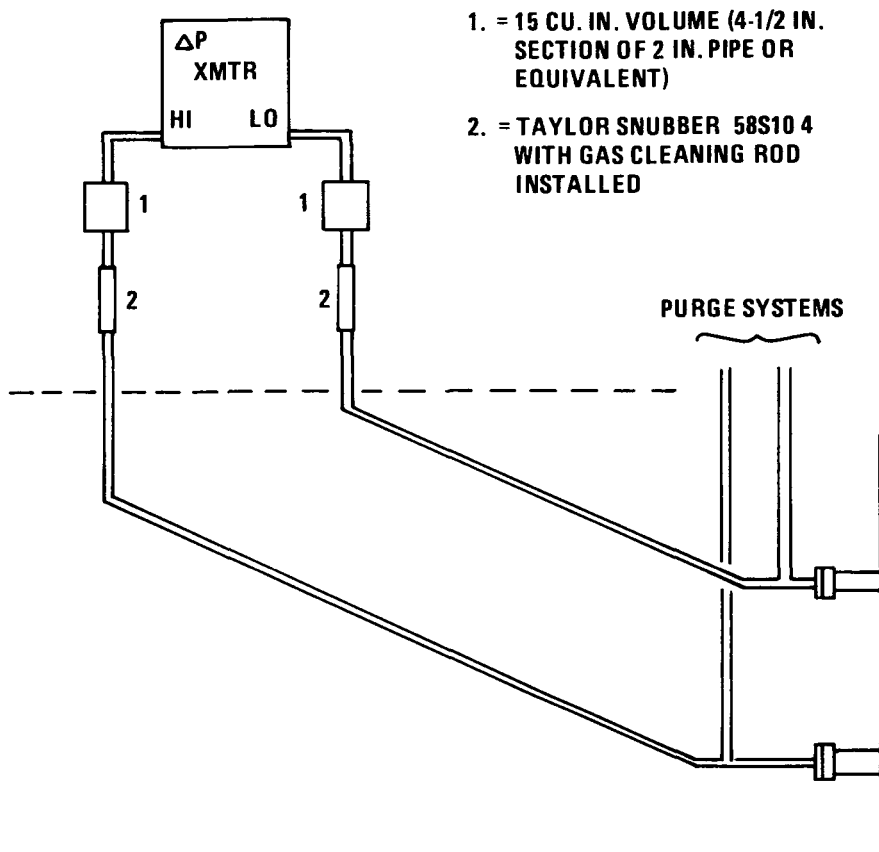


FIGURE 11.9
External damping for ΔP level measurement

Many ΔP transmitters on the market have adjustable damping in the feedback. This leaves adjustment up to the instrument mechanic, who seldom has a rational basis for setting it. But an excessive amount of damping can have an adverse effect on control; changing the damping while the process is in operation may have a serious effect on control-loop stability. This feature therefore is not recommended. Where such instruments are bought on a project, the variable damping feature should be left wide open and external properly designed, fixed damping provided where required.

Damping for Externally Mounted Displacer Transmitters

As mentioned earlier, most displacer-type transmitters have a sharp resonance in the range 1–3 cps. For typical displacer housings (“boots”), we have calculated that 10:1 attenuation at 2 cps will be obtained by installing a 1-inch-diameter orifice in the lower line connecting the displacer housing to the process vessel. This provides a fiddleproof design; the guard valves should be left wide open in normal operation and need be closed only for maintenance work. For extremely noisy installations, both internal and external damping may be used, provided the time constants are selected with some care.

It is pertinent at this point to mention certain potential problems with displacer-type transmitters:

1. If excessively long connecting lines are provided between the vessel and boot, the natural frequency of the “manometer” may be low enough to interfere with level control. This is discussed in a very interesting paper by Sanders.⁴
2. If a restricting orifice is installed as suggested in the lower line connecting the level measuring chamber to the column base (or if an internal damping chamber is used), vapor condensations may occur in the upper line. This may take place fast enough to cause a significantly higher level in the chamber than in the column base. The chamber and upper line should be insulated and heated; in difficult cases a small gas purge to the upper line is helpful.
3. As discussed later, if it is desired to have a liquid purge and damping, the purge rate must be fixed and the liquid level elevation in the chamber known. One could deliberately use the upper connecting line as a condenser, send a fixed amount of condensate via a metering orifice or capillary to the measurement chamber, and overflow the remainder back to the column.

Level Measurement Errors Due to Velocity Effect

As shown in Figure 11.10, level is sometimes measured with the nozzles located on vessels or pipelines of different diameters. Two cases are of interest: downflow and upflow.

Downflow

By far the most common case, downflow is occasionally encountered on distillation column bases or vaporizers with the upper level tap on the column base and the lower tap on the bottom-product line. If the liquid velocity V in

the bottom-product line is in feet per second, then the error h_e in inches of process liquid is:

$$h_e = -0.28 V^2$$

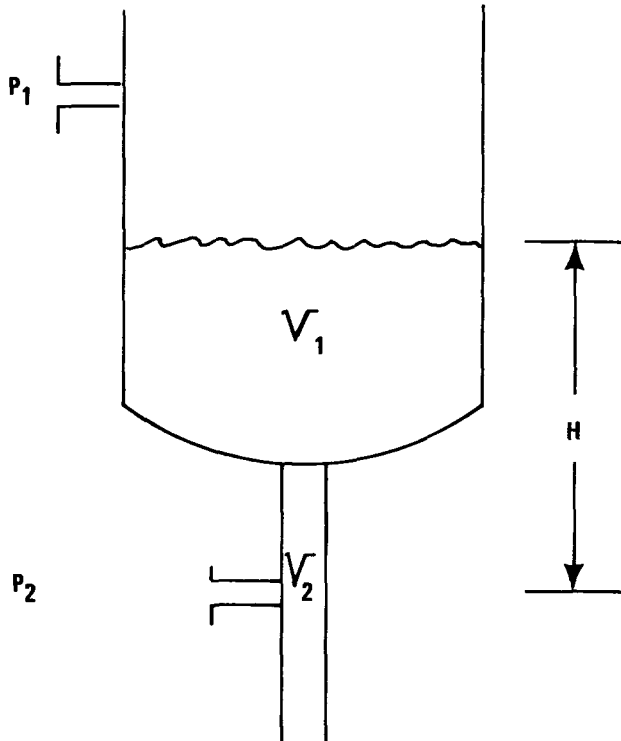
When $V = 1.9$ ft/sec, $h_e = -1$ inch. In other words, the transmitter output signal will indicate a level that is 1 inch lower than the true level.

Upflow

In this case there is no velocity error.

Specific Gravity Compensation for Level Measurements

Occasionally we find enough difference between column-feed specific gravity and normal bottom-level specific gravity that there is a serious base-level measurement error at startup. Also, at times, a given column may have a different bottom specific gravity as a result of changes in bottom-product specifications.



1. DOWNFLOW CASE

$$H_e = -0.28 U_2^2 = \text{IN. PROCESS FLUID ERROR IN MEASURING } H$$

2. UPFLOW CASE

NO ERROR

$U = \text{FT/SEC}$

FIGURE 11.10

Velocity error in head measurement

Further, if the column has an internal reboiler, a variable density froth exists above the tube bundle.³ There are many possible ways of compensating for variable specific gravity, but the one shown in Figure 11.11 was developed for a specific application.

For an understanding of this scheme, note that equation (11.52) can be rearranged:

$$H = \frac{\Delta P_H}{(g_L/g_c)\rho_L} \quad (11.56)$$

In accordance with this equation, the level transmitter signal is divided by the specific gravity transmitter signal to give a corrected level signal. As an example, consider the following:

ΔH_T = level transmitter span, feet of
process fluid

= 4.0 feet

σ_{\max} = maximum specific gravity

= 1.3

σ_{\min} = 0.8

$\bar{\sigma}$ = 1.05 = value at which level transmitter is
calibrated

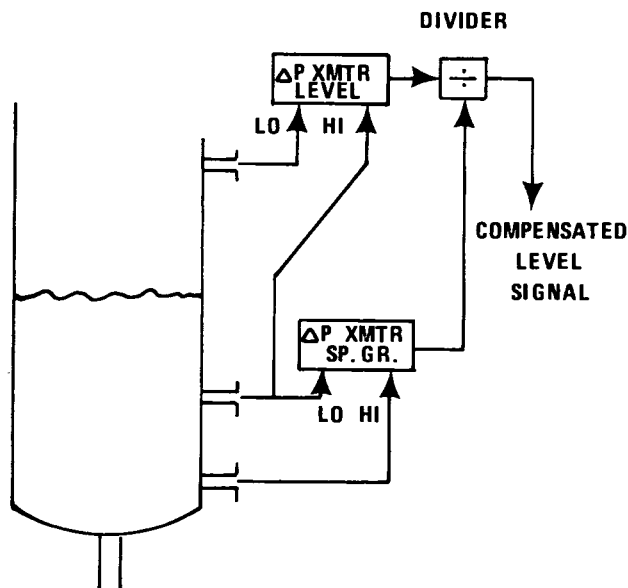


FIGURE 11.11

Specific gravity compensation of head measurement of liquid level

If, for example, a Foxboro 556 relay is used as a divider, we obtain:

$$f = \frac{1.3}{1.05} = 1.24$$

$$S = \frac{1.3 - 0.9}{1.3} = \frac{0.5}{1.3} = 0.385$$

$$Z = \frac{0.8}{1.3} = 0.615$$

where f , S , and Z are terms used by Foxboro for adjustable relay parameters.

In the absence of specific gravity compensation, it is desirable to calibrate the level transmitter so that a full-scale output (3–15 psig, 4–10 mA, etc.) is obtained for significantly less than full nozzle spacing. A common practice is to zero the transmitter at or slightly above the bottom level nozzle and to calibrate for full output at 90 percent of nozzle spacing. This permits full transmitter output if the process specific gravity is somewhat less than design. See also recommendations in Chapter 4.

Characterized Displacers

Sometimes it is necessary to operate a distillation column with the liquid level down in the bottom dish to minimize total liquid holdup. This situation usually arises because of thermal degradation. Now for level control purposes we want a transmitter whose output signal is proportional not to liquid height, but to inventory (pounds, gallons, cubic feet, etc.). To accomplish this some vendors can furnish displacers with an elliptical top-to-bottom shape. Alternatively the output of a standard transmitter may be compensated by a microprocessor or computer.

Purge System Errors

When a ΔP transmitter is used for measuring base level, it is common to locate it above the upper level nozzle and to purge at least the lower impulse line, and often both impulse lines. Generally speaking, purges are intended to accomplish either or both of two functions:

1. Sealing, which is simply the isolation of the transmitter from a clean (no solids, nonplugging) process liquid. Low-velocity laminar flow is usually adequate for sealing, but as we shall see later, may contribute to dynamic measurement errors. For these applications gas purges are usually used.
2. Flushing or sweeping, which may be required where solids in the process liquid tend to plug the impulse line. Gas purges often are not adequate. But for liquid or gas, turbulent flow through the exit diptube is recommended. This usually requires a vastly higher purge flow rate than is common practice.

Now let us look at some sources of level measurement error due to purge systems.

Unequal Purge Flows

When the level transmitter span is small, one must be careful to ensure that both purge systems are equalized—that is, that they have the same flow rate and same length of same diameter impulse or purge line. Failure to accomplish this equalization may result in an offset in transmitter reading that will not be detected by conventional calibration procedures. This consideration is particularly critical for specific gravity measurement by ΔP .

Inadequate Purge Flow

If purge rate is low enough, a rapid increase in liquid level will back liquid up into the lower impulse line. The transmitter will only gradually reflect the level change as the purge flow slowly displaces the liquid (Figure 11.12).

A rapid increase in vessel pressure can cause a different kind of error if purge rate is low enough. In this case liquid is backed up into the lower impulse line and the transmitter indicates a false low level (Figure 11.13). Again the error disappears as the purge flow displaces the liquid. It is probably apparent that either of these phenomena can play havoc with a level control system.

Static Pressure Variations

Static pressure variations on the process side cause changes in the true volumetric flow rate of gas purges. This has been particularly troublesome with vacuum columns where the change in volumetric purge flow rate from startup conditions to normal operating conditions may be 100:1 or greater. The result

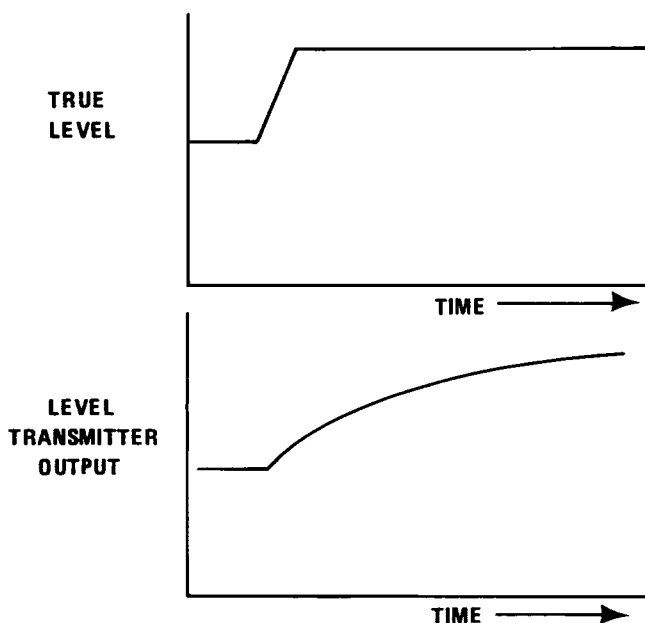


FIGURE 11.12

Insufficient purge; transmitter lag in response to rapid rise in level

may be excessive pressure drop in the impulse lines, and may even load the vacuum jets. Consequently it is recommended that gas purges not be used where process pressure is 6 psia or less unless the system is carefully designed and is calibrated after startup.

Installation Errors

Typical gas-flow purges are arranged as shown in Figure 11.14. Upstream gas pressure is high enough that critical drop exists across the needle valve, which is followed by a rotameter. The purge is then connected to the transmitter impulse line close to the transmitter. This design has a number of sources of error.

First, the rotameter does not indicate true flow since it reads correctly at only one temperature and pressure, usually 70°F and 14.7 psia. The rotameter pressure, however, will ride up and down with column-base pressure. Consequently with this design true flow is rarely known since it requires determining correct rotameter pressure and temperature and substituting both into a correction formula along with indicated flow.

Second, the fact that the purge is connected close to the transmitter rather than to the process means that there may be an error due to impulse-line pressure drop. This is rarely measured or calculated, and most common calibration procedures do not allow for it.

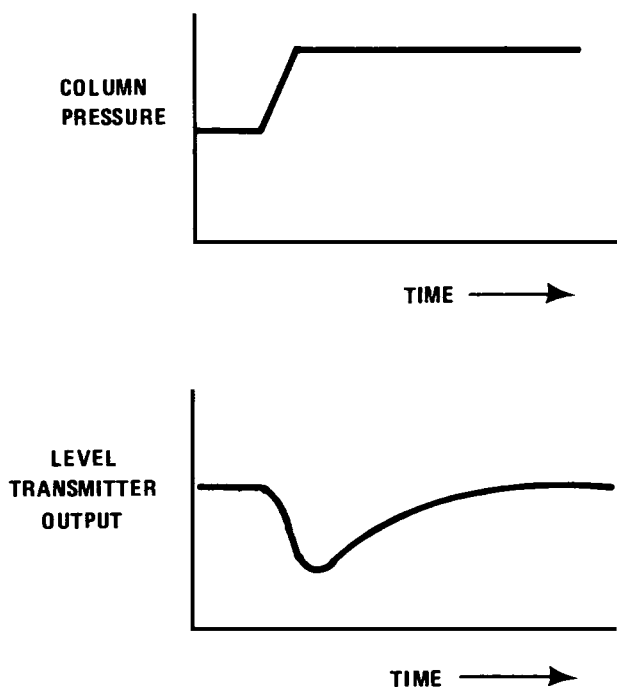


FIGURE 11.13

Insufficient purge; level transmitter erroneous response to rapid rise in pressure

Third, the restrictor usually chosen is a needle valve whose annular clearance changes with ambient temperature, particularly when plug and seat are made of different materials. Consider, for example, a needle valve with an orifice 1/8 inch in diameter. For 1.0-scfh air flow, supply at 50 psig, and critical flow, the required C_v is about 0.004. The plug-seat clearance is only about 0.0002 inch.

A far better gas-flow purge system is shown in Figure 11.15. Here the rotameter is upstream of a fixed, capillary restrictor that has been designed to provide a specific, known flow for a particular application. Further, the purge joins the impulse line close to the process rather than close to the transmitter.

An even better system, particularly for specific gravity applications, is shown in Figure 11.16. Here the fixed capillary restrictor is used in conjunction with a purge-flow regulator. In addition to being more reproducible, this system does not require critical flow.

This is probably a good point at which to comment on horizontal versus angled nozzles. For most applications the horizontal design is adequate and is cheaper. The angled design is believed by some engineers to offer advantages

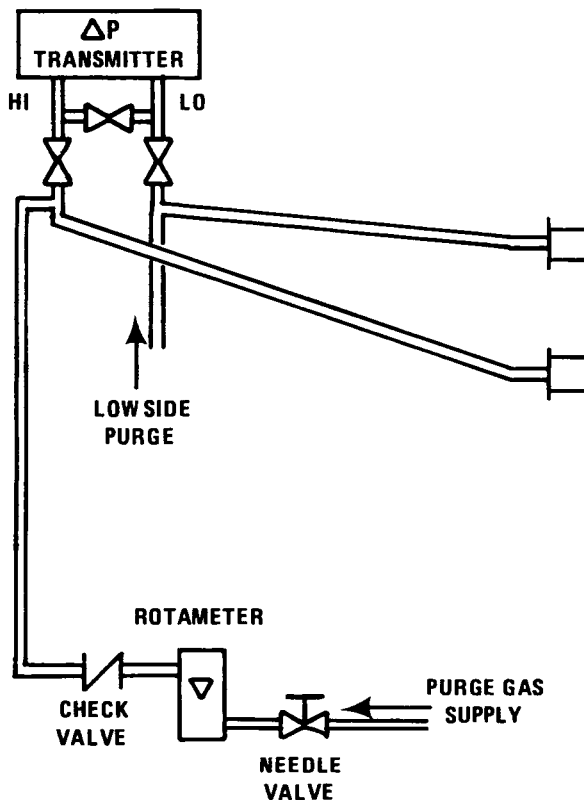


FIGURE 11.14
Typical gas flow purge system

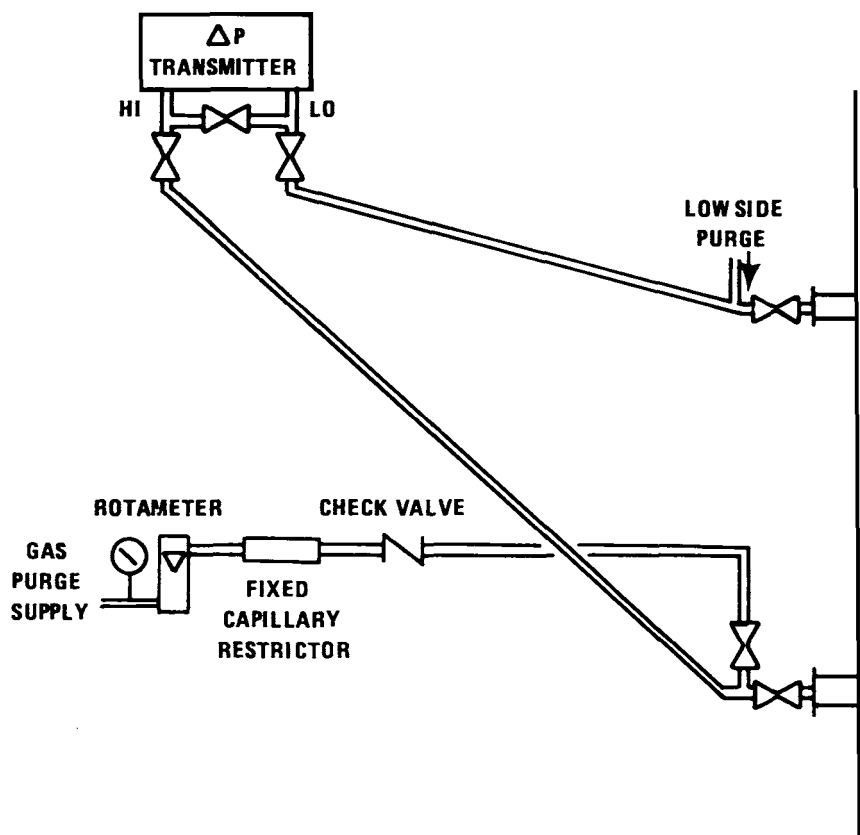


FIGURE 11.15
Improved gas flow purge system

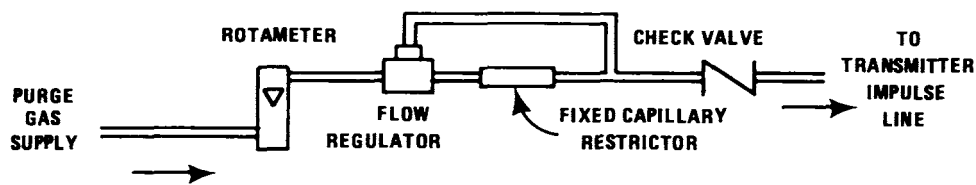


FIGURE 11.16
Best gas flow purge system

when solids are present. When used with displacer-type instruments, it reduces the available level transmitter span, particularly if a bottom connection to the measurement chamber is used.

With angled nozzles it is common practice to use a dip tube as shown in Figure 11.17. This is usually omitted with horizontal nozzles.

Liquid Purges

Liquid purges are sometimes used with ΔP transmitters for difficult applications where there is otherwise a tendency to plug the impulse lines. For accurate, constant flow, a purge-flow regulator is mandatory, and the calibration procedure should allow for impulse-line pressure drop.

Liquid purges are also sometimes used with displacer-type transmitters. Such installations should be designed to avoid any pockets in which solids can accumulate. In addition, if any damping is used, either an external orifice or an internal chamber, the calibration should allow for the extra head due to the purge flow. As an example, some years ago we had a problem where our fluid mechanics experts recommended a purge velocity of 4 ft/sec through the damping orifice at the bottom of the displacer housing to defeat plugging. We chose 3.0 gpm of a liquid with a specific gravity of 0.8. The orifice was calculated to have a diameter of 0.55 inch. The level in the housing was then 9.0 inches higher than it would have been with no purge.

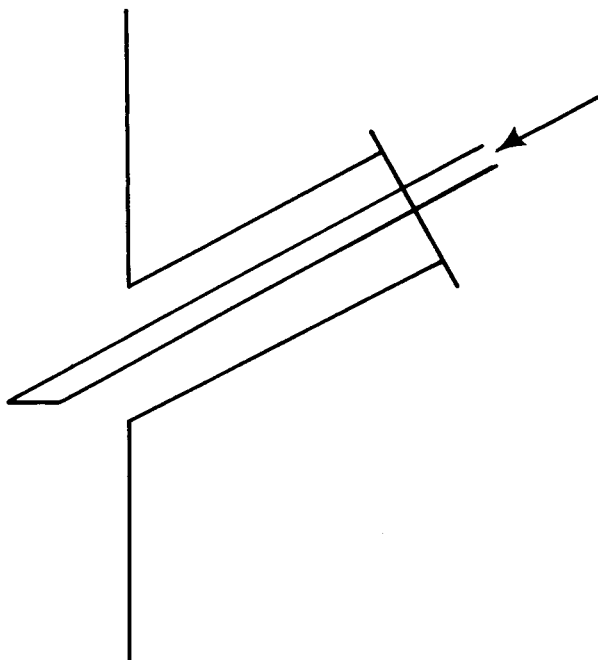


FIGURE 11.17
Angled nozzle with dip tube

Various designs for systems with unusually difficult plugging problems are discussed in a paper by Schnelle and Schmoyer.⁵

Other Head-Measurement Techniques

Other head-measurement techniques include the following.

ΔP Transmitter with Double Remote Seals (Figure 11.18)

Several transmitters of this type are now on the market. They have the advantage of providing sealing without the use of purges. If solids are present, one must use so-called extended diaphragms that are flush with the vessel interior wall. In such applications one cannot use guard or block valves. This is not necessarily a serious objection. Many processes are sufficiently hazardous that mechanics are not allowed in the area except during shutdown. In addition, some plants have experienced sufficient leakage with closed guard valves that removal of a level transmitter during operation is not permitted.

For base level measurements, internal damping chambers should be used if sealing only is required. For severe plugging applications such chambers are, of course, inappropriate.

Two Flush Diaphragm Transmitters (Figure 11.19)

Some vendors make ΔP transmitters that provide sealing only at the high-pressure connection with a so-called flush diaphragm. Sometimes such an instrument with the low-pressure connection connected to the upper level nozzle

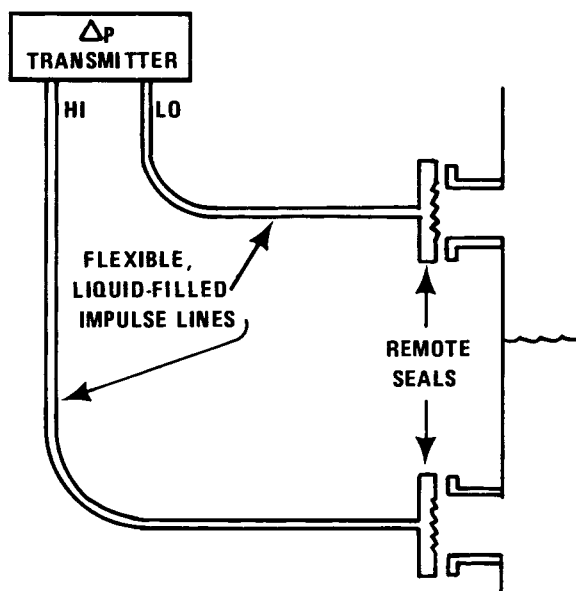


FIGURE 11.18

Level measurement with ΔP transmitter with double remote seals

is adequate. For severe applications, however, where the vapor tends to condense and crystallize or polymerize at the upper tap, two flush diaphragm transmitters are sometimes used. Their low-pressure connections are vented to atmosphere and the outputs subtracted in a summing relay.

Flush Diaphragm Transmitter and 1:1 Repeater (Figure 11.20)

A variation of the preceding uses a flush diaphragm on the lower level nozzle and a 1:1 repeater on the upper level nozzle. The output of the repeater goes to the low-pressure connection of the flush diaphragm transmitter.

11.6 CONTROL VALVES

Control valves play a very important role in continuous processes. This is especially true for distillation columns.

Typically there are four major decisions, as well as a host of minor ones, to be made about each valve:

- Valve type. What type should it be—globe, butterfly, ball, or other?
- Body size and trim size. The first refers to the pipe size to which the valve end fittings, flange or screwed connections, will mate. The second is usually specified in terms of the flow coefficient, $(C_v)_N$, for the wide-open valve.
- Inherent flow characteristic. This is the relationship between $C_v/(C_v)_N$ and $X_v/(X_v)_{\max}$, fractional lift, travel, or stroke, such as linear or equal percentage.
- Actuator and positioner. A valve must have adequate threshold sensitivity and speed of response. A piston actuator with a double-acting, two-stage, high-gain positioner provides better performance in both respects than does a spring and diaphragm actuator with or without a positioner.

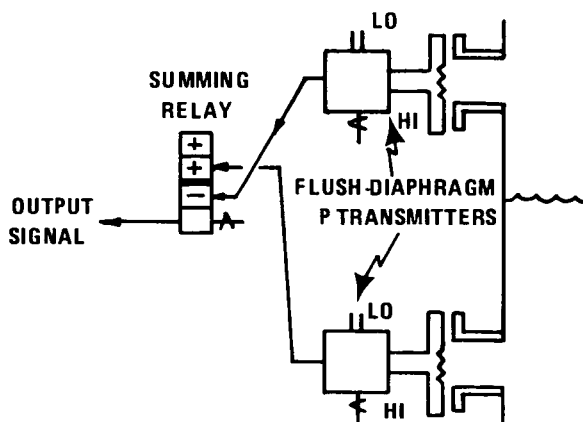


FIGURE 11.19

Level measurement with two flush diaphragm transmitters and a summing relay

Other important but non-control-related decisions include those of selecting mechanical and metallurgical features relative to corrosion, temperature, pressure, and fire safety.¹⁴

In trying to make these decisions, the engineer must recognize three significant factors that bear on control-valve selection:

1. There are today many new types of valves, including those with anti-cavitation or low-noise features. There are many disk and other rotary valves with much larger C_v (flow coefficients), lower price, and greatly reduced weight, as compared with globe valves. Some of the newer designs, however, have a greater tendency to cavitate in liquid service, or are noisier in gas service.

2. There are newer methods of calculating the normal or operating C_v that are much more accurate than the old FCI (Fluid Controls Institute) equations.⁶⁻⁸

3. In pumped systems there is a trend toward minimizing valve pressure drop to reduce pumping costs.¹⁶

General Approach

A general approach to choosing body size, trim size, and valve type has been developed that is based primarily on equations for C_v developed by an

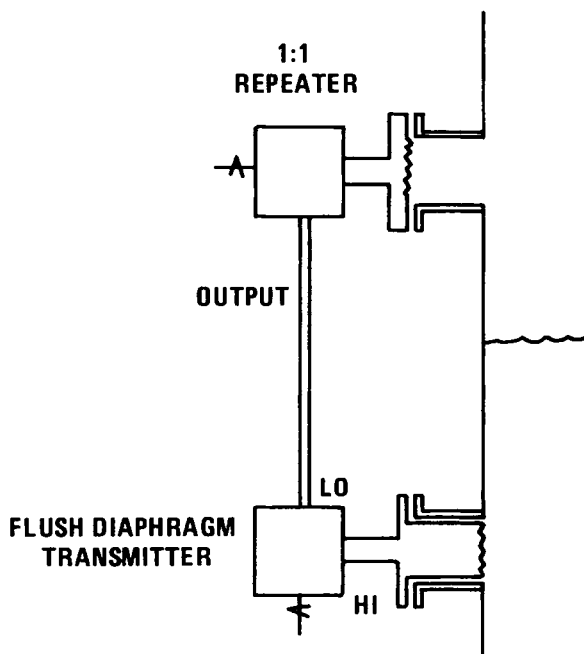


FIGURE 11.20

Level measurement with flush diaphragm ΔP transmitter and 1:1 repeater

ISA Standards Committee.⁶ These equations calculate the normal or operating C_v for the specified flow conditions. They also take into account three factors that the older equations do not:

1. Valve-body geometry via coefficient F_L .
2. The influence of associated reducers and expanders. This even permits calculating the performance of valves larger than line size.
3. Upstream liquid-vapor pressure.

To the ISA equations we have added for our own use an equation for incipient cavitation. All of this information has been combined into calculator and computer programs. The engineer first makes a trial choice of vendor, type of valve, and body and trim size. The computer or calculator program then reads out two pieces of information:

1. Operating C_v required for the flow rate and other conditions specified
2. Expected flow regimes

For liquids we have four possible flow regimes:

1. Subcritical flow, no cavitation
2. Subcritical flow, incipient cavitation
3. Critical flow, full cavitation
4. Critical flow, flashing

For gases we have subcritical and critical flow regimes.

If the program shows that the first valve choice for a liquid service will be subject to cavitation, the engineer very likely will want to look at another valve type and rerun the program. Many vendors now offer anticavitation trim, and for extreme cases, some rather exotic designs (with corresponding prices) are available.

If the valve service at original startup is to be water or some fluid other than normal process fluid, all of the previously mentioned procedures must be repeated, not only for C_v calculation purposes, but also for specification of valve operator type and size.

For gas service, if the program predicts critical flow for the engineer's first choice, a noise check almost certainly would be run to find out whether the proposed installation will conform with OSHA specifications. If it will not, various low-noise valve designs are available; the engineer may choose one and rerun the program.

With the calculated operating C_v in hand, the engineer compares it with the $(C_v)_N$ (maximum C_v of wide-open valve) for the chosen valve. At this point valve sizing, as typically practiced, ceases to be a science and becomes an art. There are no equations for calculating which size valve to purchase or install. Neither are there any generally accepted guidelines for specifying operating C_v as a fraction of $(C_v)_N$. Some engineers, for example, choose $(C_v)_N \geq 5 \times C_v$ for 50:1 equal-percentage valves. Meeting this criterion results in a valve whose travel is 60 percent or less at flowsheet conditions.

The best *inherent* flow characteristic for a given application cannot really be chosen until the desired *installed* flow characteristic, Q or w versus θ_v , is known. Here θ_v is the signal to the valve positioner. The slope of this curve is the valve gain that must be known for quantitative design of control loops. The criterion we prefer is that of choosing the installed characteristic that minimizes or eliminates the need for retuning the controller as throughput changes. For some commonly encountered systems, such as flow, flow-ratio, liquid-level, and pressure-control systems, the mathematics is simple and the results have been tabulated.¹¹

Next one must look at a particular system—pump curve, piping and equipment pressure drops, and so on—and see what inherent flow characteristic comes closest to providing the desired installed characteristic. We have calculator and computer programs to facilitate these calculations. In some cases we can help matters with a cam-operated positioner. In other cases we have installed computing relays in the control loops to generate desired curves. These curves may also be generated by the “calculated variable” function of computers and of some distributed control systems. And soon dedicated microprocessors probably will be available.

At this point it is appropriate to note that the increased complexity of the new C_v calculation procedure has led some engineers to refuse to use it except “when needed.” But the older FCI procedure offers no flags to indicate when the new procedure is required.

Maximum Flow and Turndown

A control valve *must* be able to pass a specified maximum flow and should preferably, but not necessarily, pass the minimum required flow without getting too close to the closed position. If this cannot be accomplished with one valve, a small valve that opens first may be used in parallel with a larger valve that does not start to open until the small one is wide open, or nearly so (split-range arrangement). But whether one valve is used or two, for control purposes one must be able to throttle more than flowsheet maximum flow and less than flowsheet minimum flow.

How does one determine maximum required flow? Under transient conditions a valve must be able to pass more than flowsheet maximum. But how much more? There is no simple answer to these questions. A suggested procedure is to ensure that the operating C_v calculated for flowsheet maximum flow is:

1. No greater than $0.7 [C_v]_N$ for globe valves with one-size-reduced trim. This is 91 percent lift for a 50:1 equal-percentage valve.
2. No greater than $0.5 [C_v]_N$ for valves with full-size trim for which reduced trim is not available. This is 82 percent lift for a 50:1 equal-percentage valve or 65–70° open for a typical 90° butterfly valve.

Valves sized according to either of these criteria will be less oversized than if more common sizing procedures are used (see Chapters 3 and 4).

Determining minimum required flow is even more difficult. Fortunately most wet chemical processes seldom have a turndown (maximum required flow/minimum required flow) greater than two or three to one for *normal* operation. Startups and shutdowns, however, often require very low flows for awhile. A suggested typical process turndown figure for valve sizing and selection is therefore five to one. Valve-flow turndown is defined here as the ratio of flow (*not* C_v) at 95 percent lift to that at 10 percent lift. If, for a given application, calculated valve-flow turndown is less than process turndown, two or more split-ranged valves may be required.

As an alternative to the above, it is suggested that for valve-sizing purposes we let $(Q_{\max}) \geq 1.25 (Q_{\max})_{FS}$ and $(Q_{\min}) \leq 0.75 (Q_{\min})_{FS}$. But regardless of what flow criterion is used, it is desirable that valve lift be no more than 95 percent for (Q_{\max}) and no less than 10 percent for (Q_{\min}) .

Finally, note that in a chain of process equipment, valves should be so sized that maximum manipulated flows will be greater than maximum disturbance flows. If, for example, a liquid-level control system manipulates an outlet valve whose maximum capacity is 100 gpm, the tank will overflow if inlet flow goes over 100 gpm.

Turndown and Inherent Flow Characteristics as Related to Process Applications

A. System with Pumped Liquid

For pumped systems rapidly increasing energy costs have aroused considerable interest in designing for very low valve-pressure drops. This pushes us toward line-sized valves. If piping and equipment pressure drops are large in comparison with valve-pressure drops, valve turndown suffers (becomes smaller). The installed flow characteristic also becomes very different from the inherent flow characteristic. A linear valve tends toward an installed square-root flow characteristic, while an equal-percentage valve tends toward an installed linear characteristic. At the high-flow, high-lift end of the plot, however, all curves, regardless of valve inherent characteristic, level off at a maximum flow. This is so because the valve has run out of pressure drop.

The use of small valve pressure drops makes it more difficult to get a desired installed flow characteristic than if larger valve pressure drops were used. On the positive side, small valve ΔP 's mean less tendency to flashing or cavitation.

One occasionally finds rules of thumb to the effect that valve ΔP should be at least one third or one half of system pressure drop for good control. It has been shown mathematically that higher valve ΔP 's do provide better control,² but this effect is so small for most applications that it may be neglected. A few high-speed, low-holdup chemical reactors constitute the only applications we have found to date that require high valve pressure drops. For flow to, from, and around distillation columns, valve ΔP is almost never a factor in quality of control. In view of the difficulties of accurately predicting up- and downstream pressures, however, it is probably desirable to provide at least 5 psid for the valve at flowsheet maximum flow rate.

For pumped systems where valve ΔP 's decrease with increases in flow rate, it is desirable to check the flow regime at various flow rates. A valve may be noncavitating at flowsheet maximum but cavitating severely at 50–60 percent of flowsheet maximum.

B. Process Letdown from a Higher Pressure to a Lower One

Since we will be looking increasingly at column control systems where column pressure is allowed to float, care must be taken to check valve maximum capacity at minimum ΔP_v and valve turndown at minimum flow at maximum ΔP_v . These flows, in many cases, will be manipulated to control liquid level, which means that a linear installed flow characteristic is desired. Cascading from level control to flow control is recommended.

Since, in most cases, columns with floating pressure have a higher pressure at high throughput and lower pressure at lower throughput, letdown valves will have highest ΔP_v at high flows and lowest ΔP_v at low flows. The product will usually be hot, so anticavitation trim may be required.

C. Heating-Medium Supply (Condensing Service)

The heating load here is usually a reboiler or heater. The practical problem is to determine valve downstream pressure. This requires calculating shell-side pressure of the heat exchanger as a function of heat load, pcu/hr. Line pressure drop is usually negligible. Upstream pressures and temperatures are seldom constant so orifice flow meters should be temperature and pressure compensated. For reboilers we have calculator programs that calculate downstream pressure.

D. Heating-Medium Condensate

Condensate from heat exchangers heated by steam, Dowtherm, and so forth is often collected in a pot and let down to a condensate header (or pumped away). Contrary to the case of pumped liquids discussed earlier, valve pressure drop *increases* with flow since the heat exchanger has a higher condensing temperature—and therefore a higher condensing pressure—at higher loads. This tends to make a square-root valve have nearly a linear installed characteristic and to make a linear valve have an equal-percentage installed characteristic. In most cases, however, we have found that the distortion is moderate. Note that these valves frequently must contend with cavitation or flashing.

Heating-medium required turndown is usually much greater than process required turndown.

E. Cooling Water

Water-cooled condensers and coolers are widely used in the chemical and petroleum industries. Studies have shown that cooling-water turndown requirements are usually much higher than process turndown. In fact, cooling-water turndown may be proportional to the square or cube of process turndown. Many cooling-water valves are nearly closed in the winter and wide open in summer because of the difference between minimum winter load and maximum summer load. In many applications, dual, split-ranged valves should be used.

Another vexing problem with cooling-water valves is that upstream and downstream pressures are rarely known accurately (another reason to use dual valves), except perhaps for new facilities being built at a new site. In view of this problem and the need to avoid tight shutoff in the winter, dual, split-ranged butterfly valves are recommended.

11.7 COLUMN ΔP MEASUREMENT

Column ΔP is most commonly measured by installing a ΔP transmitter above the top of the column with self-draining impulse lines. Sometimes, to facilitate maintenance, it is located above the condenser or condensers, which are usually near ground level. Vapor line ΔP , however, may introduce considerable error. The high-side impulse line is usually connected to the vapor space just under the bottom tray, and usually enters the column at an angle to facilitate draining.

This high-side impulse line is often the source of errors. It should be insulated and, in some cases, heated with steam or electricity to minimize condensation. Where this is not done, the impulse line may act as a condenser. Particularly if it is not sufficiently vertical, the line may have slugs of liquid that can cause a partial vacuum to develop. The ΔP transmitter may then indicate a sizable, negative ΔP .

Opinions vary as to the optimum high-side impulse-line size. Some engineers prefer at least a 2-inch-diameter line, while others will settle for 1 inch. We prefer to compromise at 1 1/2 inches and, if possible, to purge the line. (See discussion of purges in Section 11.4 on base-level measurement.)

It is also possible to use repeaters as discussed for level measurement in Section 11.4. This technique is not highly accurate, however, and probably should not be used for ΔP 's much under 50–75 inches of water.

11.8 TEMPERATURE-MEASUREMENT DYNAMICS

Introduction

Probably the most common method of measuring moderate temperatures of liquids and gases in the chemical and petroleum industries involves a primary element (detector) and a thermowell. The primary element is usually a thermocouple, resistance thermometer, or gas-filled expansion bulb. If a thermocouple is used, it is most apt to be in the form of a "pencil" or sheathed assembly, which provides rigidity and mechanical protection.

The instrument or control engineer is interested in the study of the dynamic behavior of thermowell/primary-element combinations for these reasons:

1. To predict the dynamic behavior quantitatively in order to design a temperature control system quantitatively.

- 2. To decide on the design of optimum thermowell/primary-element combinations.
- 3. To select optimum installation practices.

To support these interests, we have devised a mathematical second-order model that is simple enough to be used as a practical working tool for design engineers or plant engineers. The details are not reproduced here but are presented elsewhere.⁹ The model is an improved version of one discussed in Chapters 21 and 22 of reference 2. Calculator programs calculate the two time constants, τ_a and τ_b , and either the step response to ambient temperature changes or frequency response.

Illustrative Examples—Forced Convection

To illustrate the application of the model mentioned above, let us look at several cases.

Example 1: Gas Flow. Let us consider a 1/8-inch OD pencil-type thermocouple in a well 0.405-inch OD by 0.205-inch ID. The service is steam and the base case velocity is 152 ft/sec. The effect of velocity on the two time constants is shown in the following table

	$V = 300$ ft/sec	$V = 152$ ft/sec	$V = 50$ ft/sec	$V = 5.0$ ft/sec
τ_a , sec	91.8	92.2	93.4	107
τ_b , sec	5.2	7.6	14.4	49

Step-response curves are presented on Figure 11.21. If the thermocouple had been used bare, a single time constant of 1.9 seconds would have been obtained for $V = 152$ ft/sec. In view of the large annular clearance, there may be some error here in assuming a purely conductive heat-transfer mechanism for the annular fill.

For the base case of 152 ft/sec, let us examine the effect of using different annular fills:

	AIR	OIL	MERCURY
τ_a , sec	92.2	22.3	8.7
τ_b , sec	7.6	7.2	0.34

Step responses are given in Figure 11.22. Neglecting the thermal capacitance of oil and mercury as we did may introduce some error here.

Consider next the effect of changing clearance. For the same outside thermowell diameter, for $V = 152$ ft/sec, for the same thermocouple, and for air in the annular space we obtain:

	Annular Clearance		
	0.040 inch	0.020 inch	0.005 inch
τ_a , sec	92.2	52.4	16.6
τ_b , sec	7.6	8.6	8.1

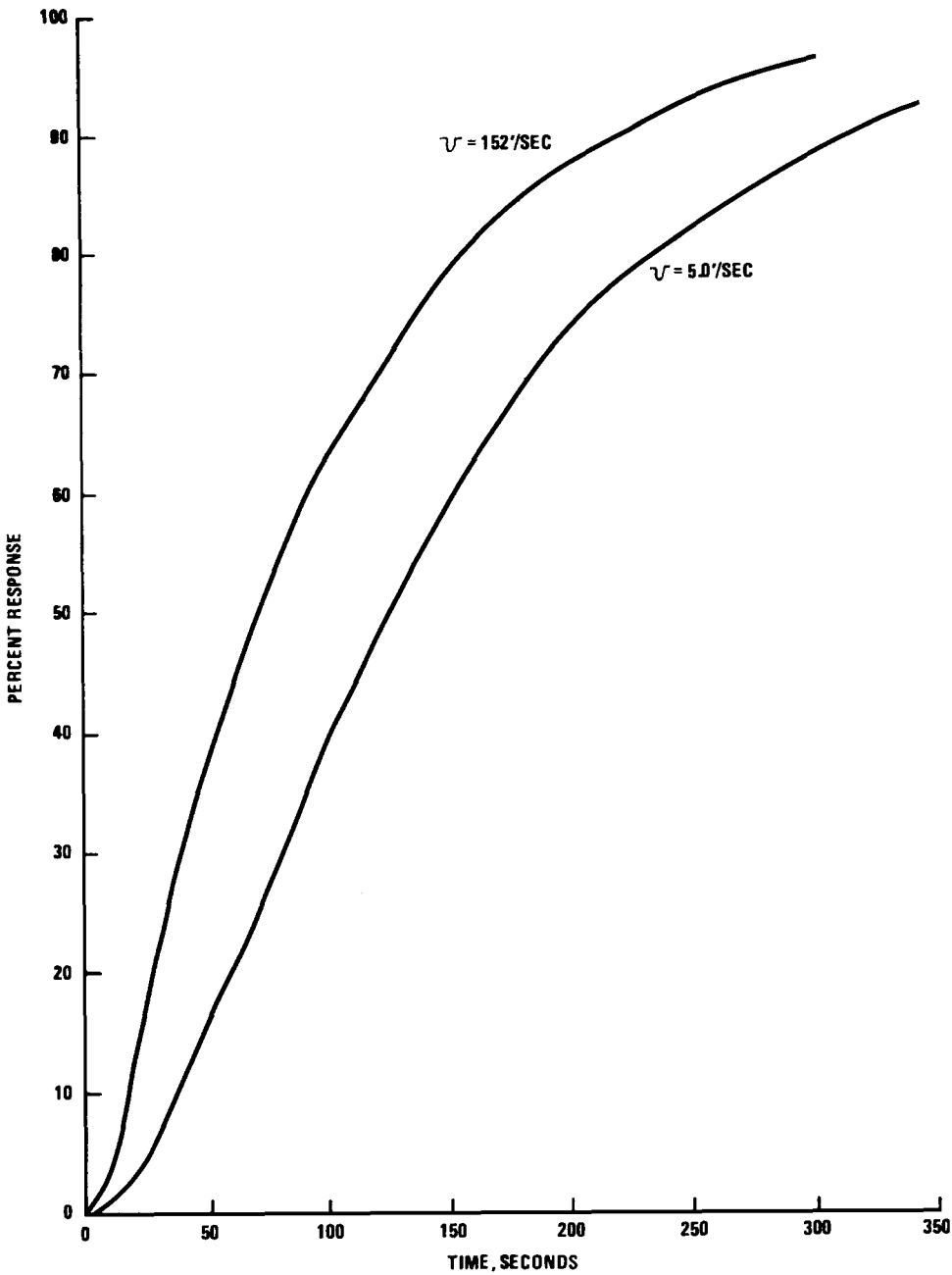


FIGURE 11.21
Effect of velocity on step response

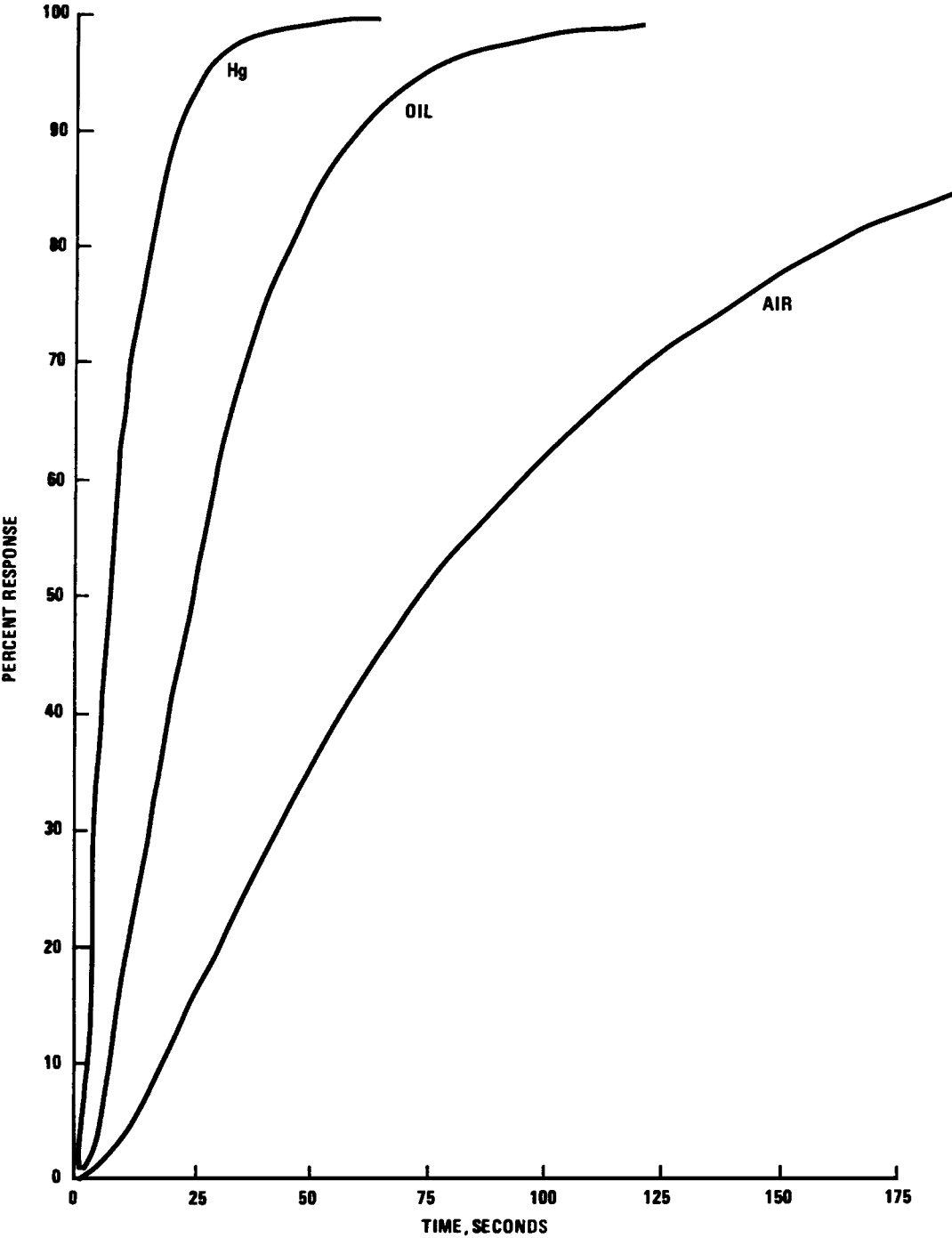


FIGURE 11.22
Effect on step response of various annular fills

As demonstrated, reducing the annular clearance from 0.040 to 0.005 inch reduces the major time constant by almost a factor of 6, even though the mass of the thermowell is increased. Comparative step responses are shown in Figure 11.23.

Example 2: Liquid Flow. Next let us consider flow of an organic liquid. Thermowell has a 0.525-inch OD by 0.260-inch ID. The pencil-type thermocouple has a 0.250-inch OD. Let us look first at the effect of different liquid-flow velocities.

	$V = 10.0$ ft/sec	$V = 1.0$ ft/sec	$V = 0.1$ ft/sec	$V = 0.01$ ft/sec
τ_a , sec	25.3	26.3	31.6	61.5
τ_b , sec	1.6	4.1	10.3	16.5

In this particular case, the major time constant changed very little until velocity became very small. Comparative step responses are shown in Figure 11.24. If the thermocouple had been used bare in the first case (10 ft/sec), a single time constant of 0.52 second would have been obtained.

Next let us consider the effect of different annular fills: air, oil, and mercury, all for 10-ft/sec velocity.

	AIR	OIL	MERCURY
τ_a , sec	25.3	6.53	2.25
τ_b , sec	1.7	1.51	0.19

For oil the value of thermal conductivity used was $0.079 \frac{\text{pcu/hr}}{\text{ft}^2(^{\circ}\text{C}/\text{ft})}$. Step responses are plotted in Figure 11.25.

Consider next the effect of varying thermowell internal diameter while holding the external diameter constant. Again, in this example velocity was maintained at 10 ft/sec and annular fill was air.

	Annular Clearance		
	0.055 inch	0.005 inch	0.0005 inch
τ_a , sec	228	25.3	3.8
τ_b , sec	1.0	1.6	1.3

Step responses are given in Figure 11.26.

Applications

For large-scale processes, one often hears the arguments that fast temperature measurement is not important because large distillation columns, heat exchangers, and so on are slow. In reality the size of such equipment has little to do with its dynamics, which are determined mostly by the service. For instance, reboilers are very fast, condensers a little less so, liquid-liquid heat exchangers slower yet, and gas-gas exchangers slowest of all—but we frequently find cases in which temperature measurement provides the major lag or lags in a control loop.

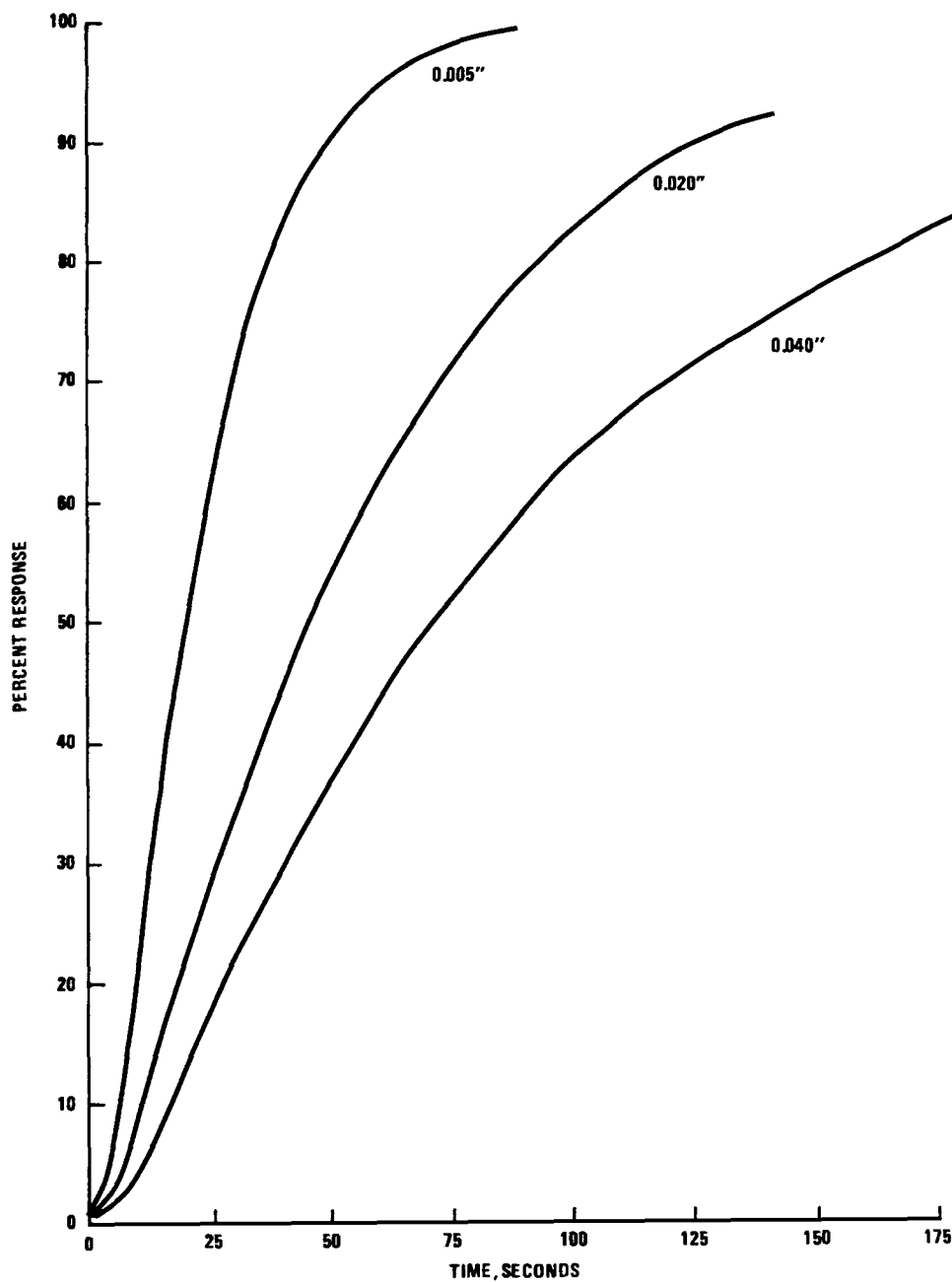


FIGURE 11.23
Effect on step response of annular clearance

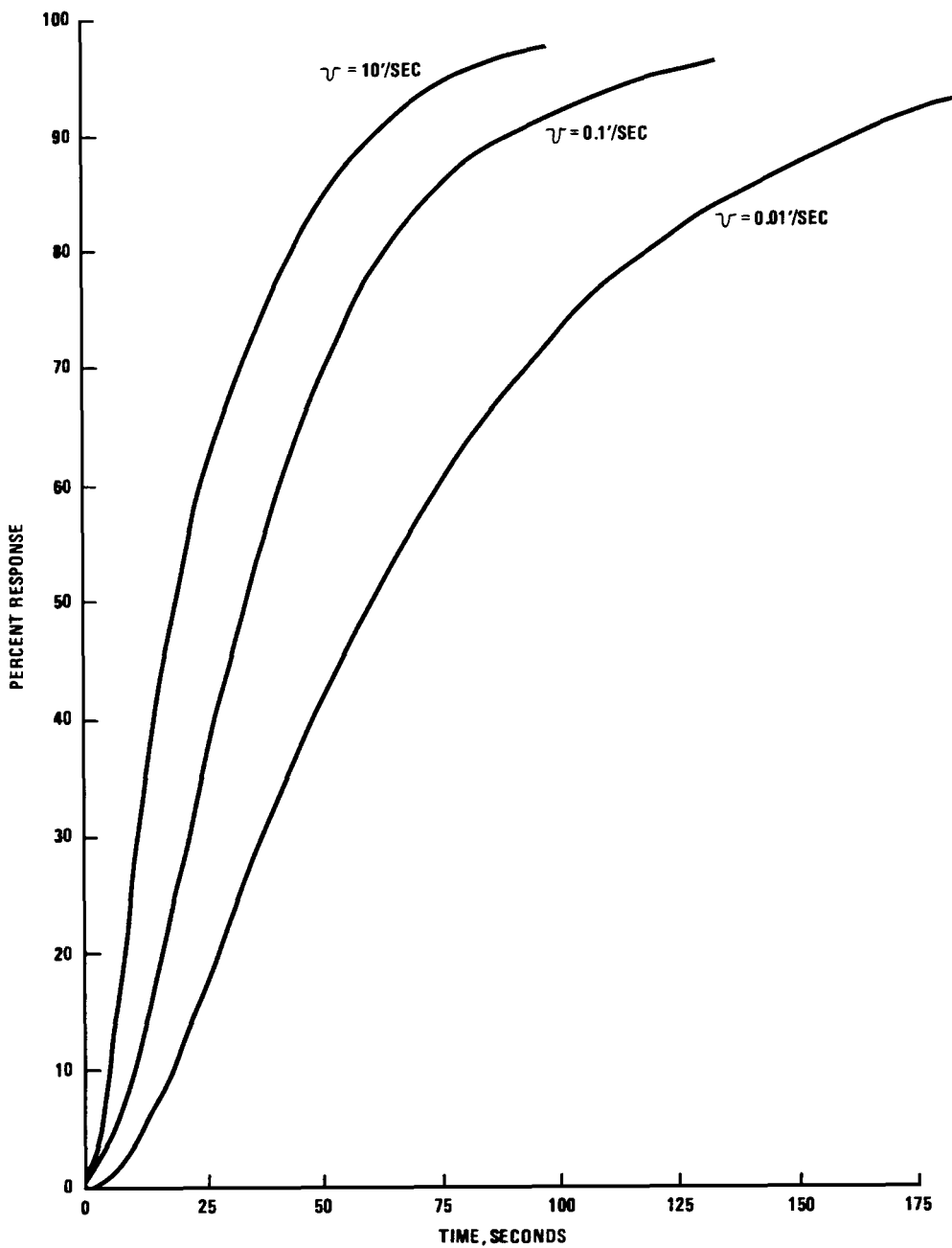


FIGURE 11.24
Effect of fluid velocity on step response

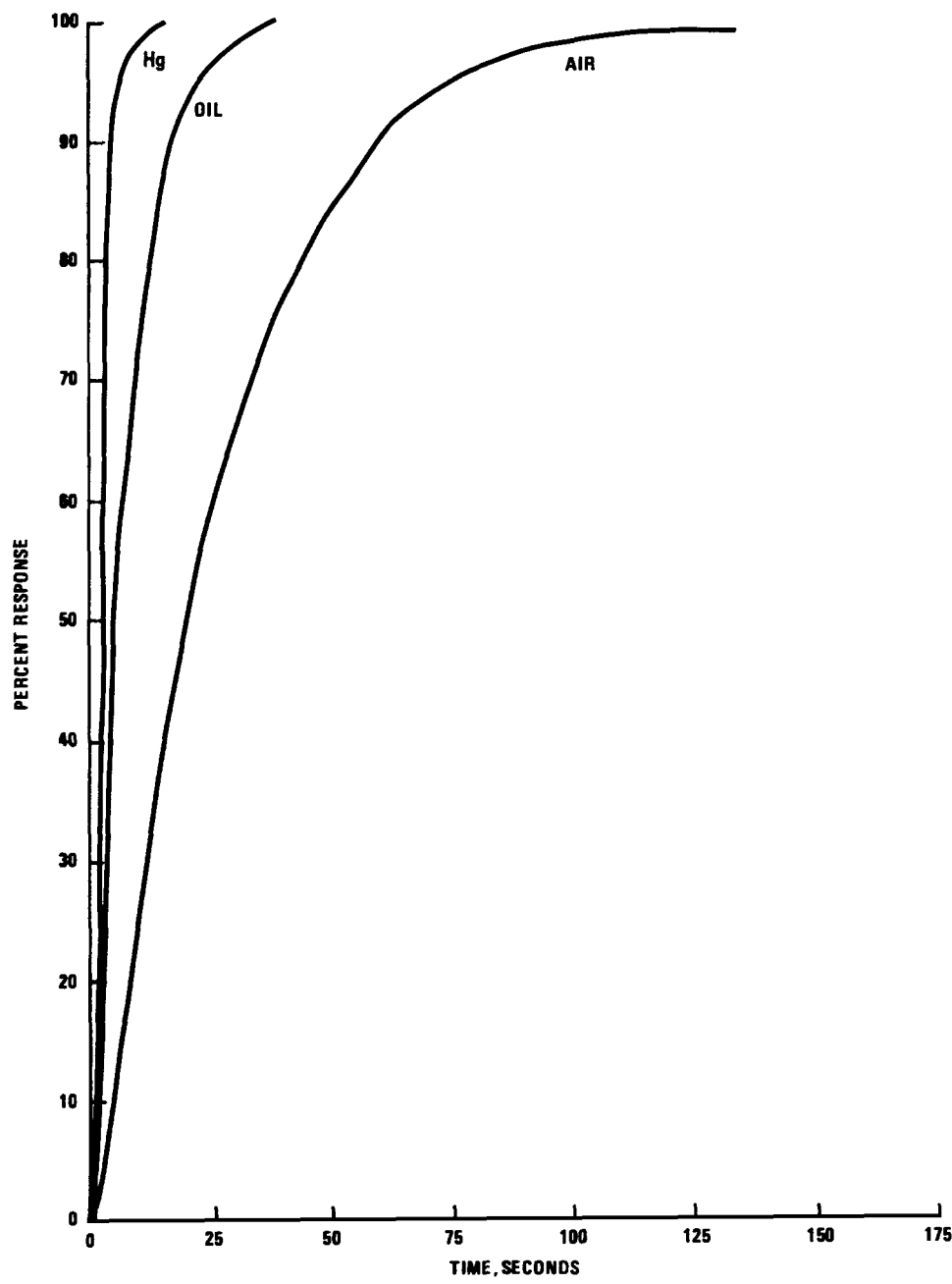


FIGURE 11.25
Effect of annular fill on step response; $v = 10/\text{sec}$

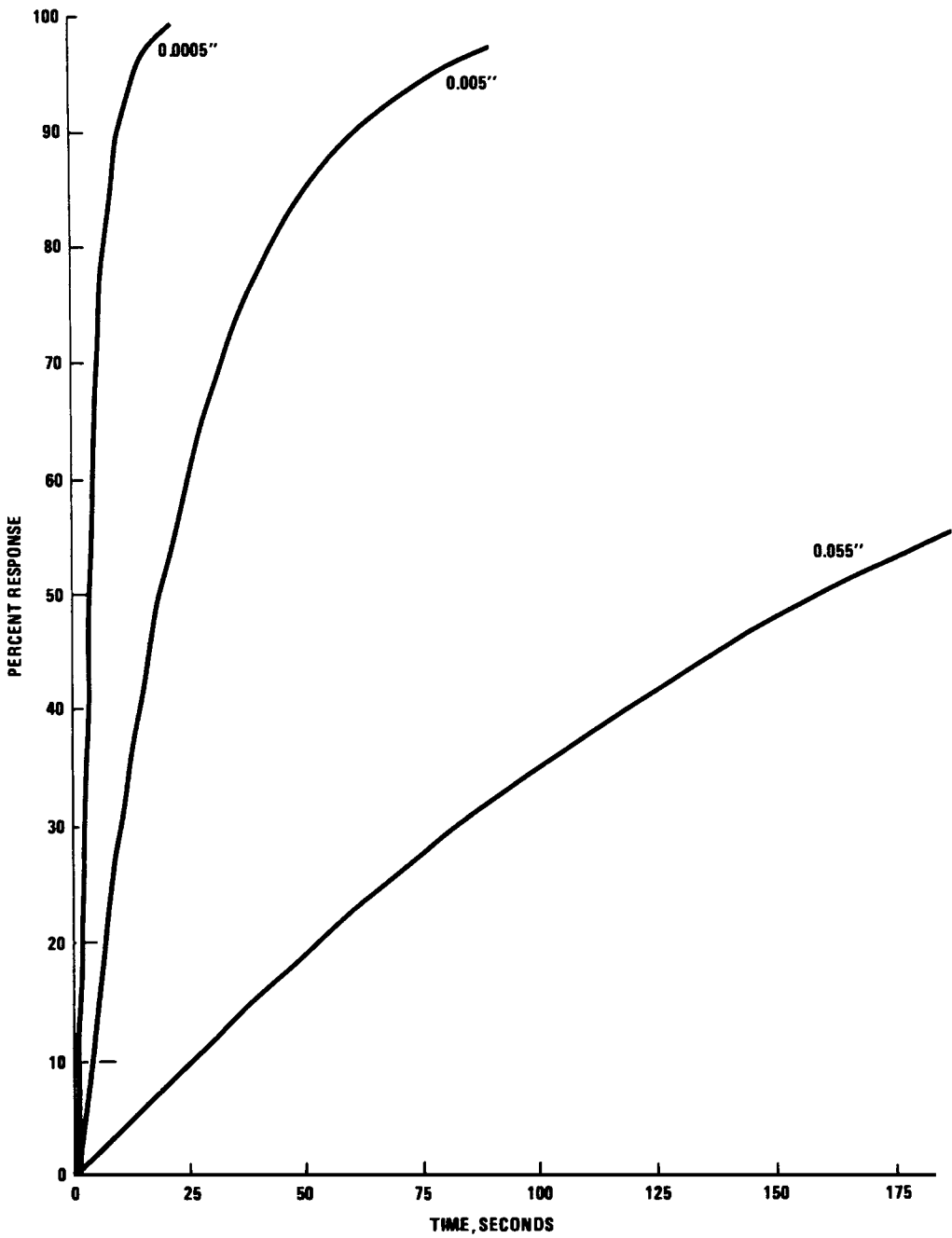


FIGURE 11.26
Effect of annular clearance on step response

To obtain a good temperature measurement, the following guidelines are suggested:

1. Use a bare detector if possible—many thermocouples, resistance thermometers, and expansion bulbs may be used without thermowells.
2. If the foregoing is not feasible, choose a minimum-sized thermowell-detector design in which annual clearance is 0.005 inch or less.
3. Provide an annular fill of oil, mercury, or pleated aluminum foil.
4. Install the thermowell in a forced-convection environment; avoid natural convection environments.
5. If possible, install the thermowell parallel to the direction of flow and pointing upstream.
6. If exploratory calculations show that film resistance is too high, install the thermowell in a section of pipe with reduced diameter to increase fluid velocity.

11.9 FLOW AND FLOW-RATIO CONVENTIONS

Linear Flow Measurement Convention

A simple but very important instrument convention that should be followed routinely is that of using linear flow measurements. If orifice flow meters are used, they should be followed by square root extractors. There are three reasons for this convention:

1. If a square-law flow meter is used, it can be shown¹¹ that the most desirable installed valve characteristic is also square law. This is to maintain constant stability as flow rate changes. If a linear flow meter is used, the optimum installed valve characteristic is linear. Most instrument engineers choose equal-percentage valves whose installed characteristics most often are somewhere between equal percentage and linear. A linear flow measurement is therefore a better choice.
2. Flow-control loops are often secondary loops in cascade systems. Linear flow measurements are a must since square-law flow measurements would cause the primary controllers to become unstable at low flows. This is so because the flow loop gain, as seen by the primary controller, approaches infinity as flow approaches zero. With linear flow measurements, this problem does not exist.
3. If flows are to be added or subtracted, flow measurements must be linear.

Flow-Ratio Conventions

Two more important conventions relate to flow-ratio control:

1. The divider technique, whereby a manipulated flow is divided by a wild flow, is preferable to the multiplier technique. With the latter a wild flow is

multiplied by a gain factor (sometimes remotely set) to calculate the set point for a manipulated flow controller. The former lends itself well to our preferred antireset windup scheme for cascade control where the latter does not. See item 2, however.

Although the divider technique is sometimes objected to on the grounds that the flow-ratio loop has nonlinear loop gain, it has been shown^{2,12} that both techniques are characterized by nonlinear gain. With linear flow measurements, the divider causes less nonlinearity than if square-law measurements are used.¹²

2. Cascading temperature or other variables to flow-ratio control should be avoided unless ratio turndown and flow turndown are less than 2:1. This applies whether one uses the divider (closed-loop) or multiplier (open-loop) flow-ratio scheme. The problem is that the gain of the ratio loop as seen by the primary loop is zero at low flow and high at high flows.¹² We have not found a convenient way to compensate for this with analog equipment without interfering with antireset windup.

An alternative that works well is to cascade to flow control and to use our preferred impulse feedforward scheme (see Chapter 12) for the wild flow.

11.10 CONTROL-VALVE SPLIT RANGING

“Split ranging” is a term applied to techniques for controlling two or more valves from one controller. If overrides are involved, it may mean controlling two or more valves from two or more controllers. In past years split ranging was accomplished by installing special springs in valve positioners. This technique is not very flexible, however, and plant maintenance people do not like to have to keep track of nonstandard valves.

A more flexible approach that has gained wide acceptance is that of using two or more amplifiers between the controller or override outputs and the valves. In pneumatics the amplifiers are commonly fixed-gain relays with adjustable bias. A simple dual-valve installation is shown in Figure 11.27. Here we have a large valve and a small one in parallel. The small one opens first; the large one remains closed until the small one is fully open. Such an arrangement is used to get a large flow turndown. For the small valve, the gain 2 relay is biased such that a 6.0-psig input signal causes the output to be 9.0 psig. In effect this bias makes the small valve signal range 3.0–9.0 psig. The gain 2 relay for the large valve is biased such that a 12.0-psig input signal causes a 9.0-psig output. In effect this makes the large valve signal range 9.0–15.0 psig.

This technique, as mentioned earlier, has great flexibility. For example, the relays may be so biased that the large valve starts to open slightly before the small valve is wide open. Or the relays may be biased so that the large valve does not start to open until well after the small valve is wide open.

As another example, consider Figure 11.28. Normal level control of the reflux drum is via the distillate valve. The level controller is a gain 2 relay biased such that a 9.0-psig signal produces an output of 9.0 psig. This means that if

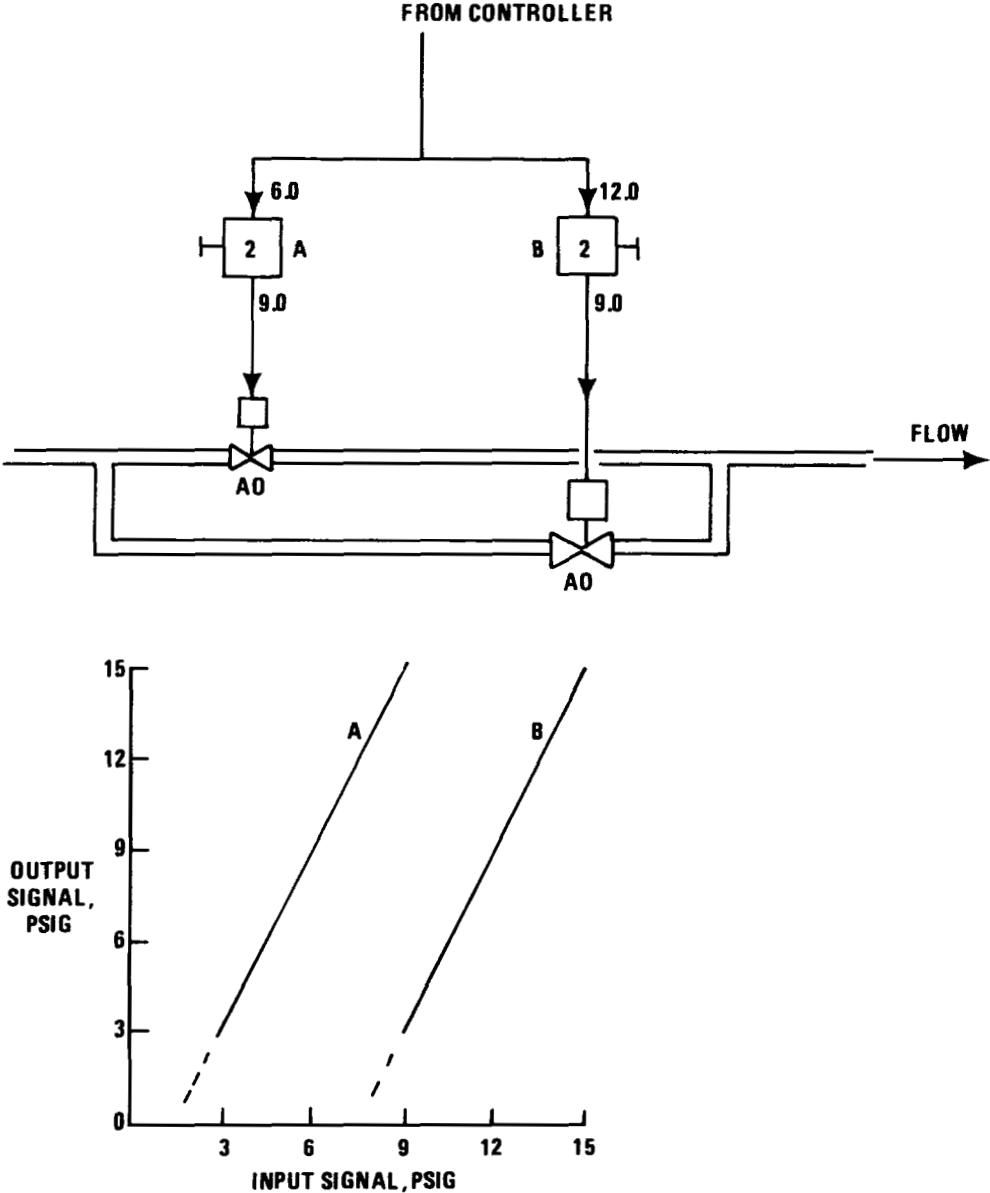


FIGURE 11.27
Split range of large and small valves

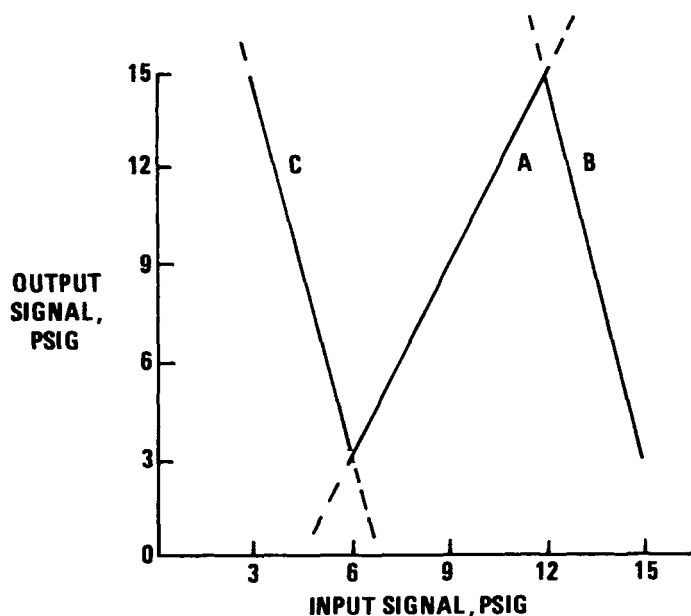
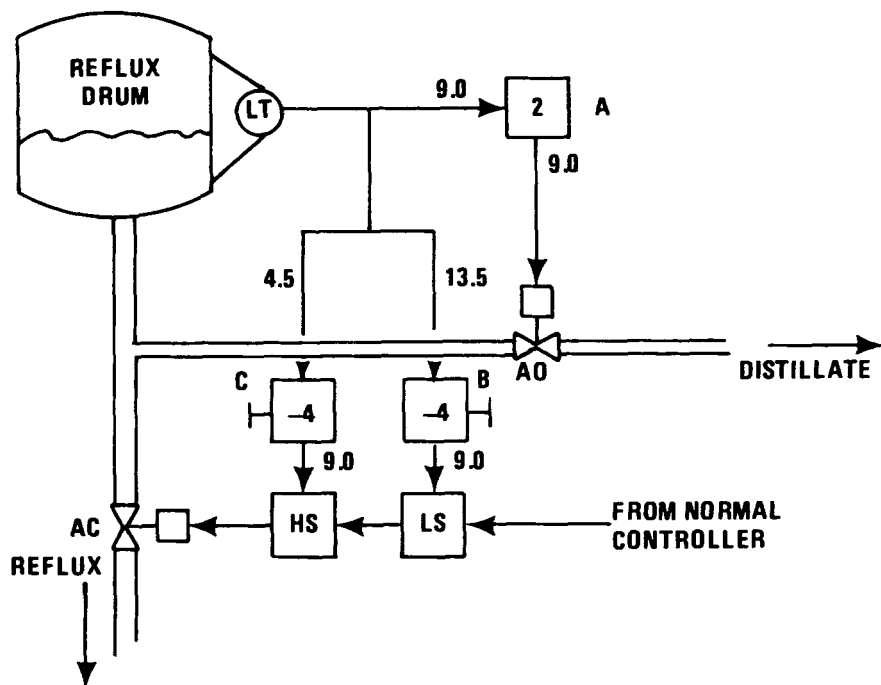


FIGURE 11.28
Split ranging reflux and distillate valves

the level drops to 25 percent, the distillate valve is closed; if the level rises to 75 percent, the distillate valve is wide open.

Suppose now that the level rises above 75 percent or drops below 25 percent. To protect against high level, we provide a gain -4 relay so biased that at 13.5 psig from the level transmitter, the relay output is 9.0 psig. This gives the air-to-close reflux valve an effective signal range of 12.0–15.0 psig in terms of liquid level. To protect against low level, we provide another -4 relay so biased that for a level transmitter signal of 4.5 psig, the relay output is 9.0 psig. This relay makes the effective signal range of the reflux valve 3.0–6.0 psig in terms of the level transmitter.

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12

Approaches to Quantitative Design

as mentioned earlier, it is our intention to provide, where possible, quantitative procedures for the design of multivariable distillation control systems. How do we go about this? Today there are four basic approaches to control system design,* three of which are quantitative.

12.1 WAYS OF DESIGNING CONTROL SYSTEMS

Instrumentation

As discussed in Chapter 1, this is a qualitative approach based primarily on past practice and intuition. The end is a multiplicity of single-loop controls. This approach, salted with a modest understanding of control theory, often provides a good starting point for quantitative design.

Transfer Function

Before the age of computers, electrical engineers adapted perturbation techniques originally worked out by physicists to develop a body of control theory based on linear differential equations. Frequency response, signal flow diagrams, and Laplace transforms are the primary working tools. Although originally intended for single-input, single-output (SISO) loops, this approach has been extended to moderately complex noninteracting, multivariable systems. It is almost the only available analytical (noncomputational) approach to control system design. Today it is the cornerstone of useful process control theory.

* It is assumed from here on that the reader is familiar with differential equations, Laplace transforms, z-transforms, frequency response, and other elements of SISO (single-input, single-output) control theory.

Simulation

The availability of computers beginning in about 1950 permitted two advances: (1) the ability to work with nonlinear differential equations (usually ordinary), and (2) the ability to study larger systems. Analog computers were used at first but have been displaced mostly by digital computers.

The first card-programmable pocket calculator, the HP-65, appeared on the market in 1974. Rapid evolution has taken place, and today there are many small but powerful machines that can be purchased for about the price of a color television set. These permit the engineer to design many commonly encountered control systems while seated at a desk. For the HP-97 and HP-41C, we have a large collection of control system programs.

Small but powerful interactive computers with CRTs and printers, usually programmed in BASIC, are also available. The HP-9845B is a good example. It can be used as a “smart” terminal to access mainframe computers. In addition, we have terminals for minicomputers that permit large, tray-to-tray dynamic simulations coded in FORTRAN or Pascal. A smaller BASIC computer, the HP-85, is proving to be a very useful portable machine.

Use of both the transfer-function and simulation approach has been very fruitful. The former can be employed to establish overall control concepts and structure, while the latter permits including nonlinearities as well as prediction of overall system performance.

Newer Methods

Since about 1960 a variety of newer mathematical and computational techniques have been introduced. These are sometimes called, collectively, “modern control theory”.^{*} Their application to industrial process control so far has been limited. This is partly so because few practitioners are familiar with the technology, and partly because it is not really appropriate for many applications. Study of control systems by these methods requires a fairly capable computer; on-line implementation usually also requires an on-line computer. Two of the best known approaches are the (1) multivariable frequency-response method, pioneered by Rosenbrock and associates in England,¹ and (2) the “state space” or “state variable” approach, which has been popular for aerospace and military applications.

In the last few years, the process control theoreticians studying multivariable control have been focusing on (1) adaptive control and (2) predictive control. Adaptive control is geared to nonlinear processes in which one set of tuning parameters may not suffice. Distillation composition is nonlinear, although not very severe compared with waste neutralization, for example. There are many adaptive techniques, ranging from simple PID controllers that adjust the tuning according to a predetermined relationship between a measured variable and controller tuning parameters (“gain scheduling”) to statistical types where an internal model is statistically fitted to the measured response and the tuning

^{*} One author can recall this term being applied to the Ziegler–Nichols papers of the early 1940s, and again about 1950 to transfer-function techniques.

parameters are determined to achieve some prescribed criterion of performance such as minimum variance. The reader is referred to Bristol,²¹ Åström,²⁰ and Sastry,²² to name a few, for further explanations.

Predictive control describes control algorithms that use an internal discrete convolution to predict the process output, and then adjust the controller parameters so that the process follows a specified trajectory. This approach is suited to processes that are difficult to model parametrically, are multivariable, and have constraints. The reader is referred to Cutler,²³ Hoopes,²⁶ Mehra et al.,²⁴ Touchstone et al.,²⁷ and Morari²⁵ for details.

Some useful insights are provided in a recent book by McAvoy²⁸ and a paper by Ray.²⁹ An interesting industrial perspective is presented in a paper by Doss et al.³⁰

There seems to be little incentive to apply this newer technology to most flow, liquid-level, and pressure controls, but one can visualize a number of other process control situations where its use would be beneficial:

- Interacting systems with more than two inputs and two outputs.
- Interacting systems with two inputs and two outputs where straight decoupling (noninteracting design) is not satisfactory.
- Systems with time-varying parameters requiring adaptive tuning. pH control systems with variable buffering are an example. This logically leads to consideration of self-tuning regulators such as discussed by Åström.²⁰
- Systems where time-optimal control is economically attractive. Batch reactors and batch distillation are examples.
- Systems where controlled variables cannot be measured directly but must be estimated from other measurements (see Chapter 10, references to work of Brosilow et al.).

Since there are real economic incentives for more extensive use of multivariable control systems, and since it is becoming increasingly common to have on-line computers, it is likely that these techniques will be used more in process control in the future.^{2-4,18,19}

12.2 KINDS OF INFORMATION AVAILABLE

In almost all design projects, there is not enough time, money, or personnel (especially qualified personnel) to do as good a job as we would like. There is, therefore, great pressure to make the necessary design decisions as quickly and inexpensively as possible. Consequently it is common practice to start with the simplest possible methods, and to depart from them only when necessary. Those rare engineers who have extensive training in process control and who are also experienced have a wide selection of tools with which to work.

In the following categories, we progress from the most basic kind of information, which permits the quickest decisions, to more sophisticated knowledge. The latter requires increasing understanding of control theory and more calculations.

Practices

There exists in the literature, and in the heads of experienced engineers, a body of knowledge about good practices: process configurations, hardware installation, control system structure, and so on. Part II of this book tries to assemble this kind of information for distillation columns.

Process Control Catalog

Here we begin to use quantitative procedures. We find, for example, simple algebraic formulas for selecting level and pressure controller settings. We also have tuning charts for flow and flow-ratio controllers.

Control Loop Catalog

Here we have a collection of open-loop transfer functions with simple equations, charts, or tables for finding controller gain, reset time, and closed-loop resonant frequency. An example is:

$$\frac{K_L e^{-\alpha\tau s}}{(\tau s + 1)(\beta\tau s + 1)}$$

The engineer who knows that this transfer function fits a given situation can quickly find controller parameters with a minimum amount of calculation and a minimum knowledge of theory.

Catalog of Open-Loop Process Transfer Functions

At this level simplicity begins to fade. Here we have transfer functions for pipelines, heat exchangers, and distillation columns, for example, as well as instruments. Since two or more of these transfer functions may have to be combined, and since there are many possible control systems, there are no simple relationships for controller tuning. Instead the engineer must use analytical methods, including frequency response, or simulation methods. Since, however, deriving process transfer functions is often tedious or difficult, or both, the catalog is usually a substantial time saver. We have a sizable collection of calculator (HP41-C) and computer (HP-85) programs, including one that takes an open-loop transfer function and automatically calculates loop gain for a specified M_p , resonance peak magnification for the function:

$$\frac{KG(j\omega)}{1 + KG(j\omega)}$$

Instrument Dynamics Catalog

Although not assembled in one place, there exists considerable information today on the dynamic performance of transmitters, valves, controllers, relays, and so forth. This is usually in the form of frequency response or step response.

For some applications, such as temperature measurement, we have HP-41C programs that calculate either step or frequency response.

“Book” Design of Control Loops

When the above-mentioned sources are not adequate, one must resort to the textbook approach: Start from the beginning, write the equations for all elements in the loop, and use frequency response or simulation or one of the newer methods to design the control loop. Today, with over 35 years’ experience in designing control systems by quantitative methods, we resort only occasionally to this method.

If all else fails, it may be necessary to determine process or instrument dynamics, or both, experimentally. In a recent book, Hougen⁵ presented an excellent discussion of techniques and equipment.

12.3 FUNCTIONAL LAYOUT OF CONTROL LOOPS

The overall functional layout of a control loop is extremely important. The design must accommodate such diverse requirements as:

- Minimum operator attention for startup and shutdown, and for manual–automatic switching
- Anti reset-windup for loops with overrides (also simplifying manual–automatic switching)
- Feedforward and interaction compensation without interfering with either normal reset or anti reset-windup
- Use of Smith predictor and other advanced controls without interfering with either normal reset or anti reset-windup

Satisfying these requirements is not basically a matter of either hardware or software. Rather it is a function of algorithms (mathematical functions) and their topological relationship to one another in a control-loop layout. Many commercial controls, whether analog or digital, are seriously deficient in one or more of the requirements listed previously. As of 1981 most vendor-designed digital systems were less flexible and less versatile than the best analog systems, but there were indications that this might be changing.⁶

A single loop with external reset feedback (i.e., feedback from the valve loading signal, not from the controller output), and with optional overrides, predictor, and feedforward compensation (or decouplers), is shown in Figure 12.1. Few plant loops have all of these features, but the point is that if a PI control algorithm with external reset feedback is used, they all fit together in a simple, straightforward fashion. This structure satisfies all of the design requirements stated earlier. It also easily accommodates fully automatic startup and shutdown—only a switch and a lag unit need be added. Note that the controller gain, K_c , and the reset time, τ_R , may be nonlinear or may be adaptively

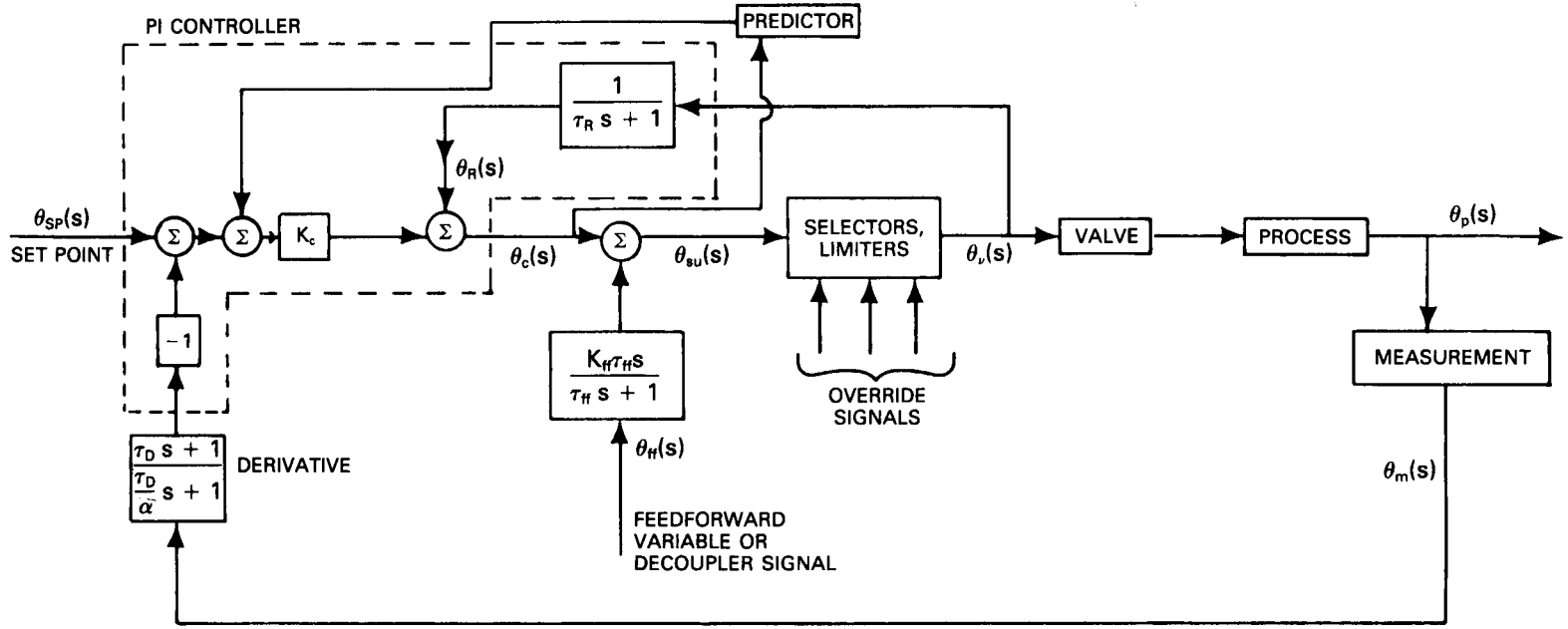


FIGURE 12.1
Single loop with feedforward compensation, derivative, PI controller, overrides, and predictor

set from external signals. The configuration shown is for “increase–decrease” controller action; for “increase–increase” action, the -1 box shown in the measured variable channel would be moved to the set-point channel.

A single feedforward function is shown but, as will be indicated later, additional functions may be required for more complex control systems. This feedforward function is also shown ahead of overrides and limiters, which might not always be the case.

It may be noted that Laplace transform symbolism has been used in this and the next two illustrations to represent reset, derivative, and feedforward functions. Although not totally valid for digital computers or microprocessors, it is nearly so when high sampling rates are used. It is not valid for nonlinear functions, and either difference equations or z -transforms are better suited for some digital applications. The basic structure of Figure 12.1, however, is being increasingly accepted by control engineers.

The theory and mathematics involved in combining feedforward compensation, overrides, controllers (PI and PID), and anti reset-windup have been covered in detail elsewhere⁷ and are reviewed briefly below. At this point we wish to point out that (1) the feedforward function, $K_{ff}\tau_{ff}s/(\tau_{ff}s + 1)$, commonly termed “impulse feedforward,” is a convenient way of feeding forward without interfering with reset when external reset feedback is used, and (2) making the feedforward time constant, τ_{ff} , equal to the reset time constant, τ_R , is usually desirable. This approach to feedforward also provides a convenient way to connect interaction compensators (decouplers) into control loops.

Note that τ_R is the reset time constant, τ_D is the derivative time constant, and α is a constant, usually in the range 6–20. Derivative is shown here as a separate unit, but is sometimes combined with PI in analog devices. Care should be taken to avoid putting derivative inside the reset loop.

Following is a brief review of the mathematics of a PI controller with external reset feedback and impulse feedforward compensation. The basic PI controller equations are:

$$\theta_c(s) = K_c[\theta_p(s) - \theta_m(s)] + \theta_R(s) \quad (12.1)$$

and

$$\theta_R(s) = \frac{\theta_v(s)}{\tau_R s + 1} \quad (12.2)$$

If there are no overrides and no feedforward signal, $\theta_v(s) = \theta_c(s)$ and we can combine equations (12.1) and (12.2) to get the conventional PI controller equation:

$$\theta_c(s) = \theta_v(s) = \frac{K_c(\tau_R s + 1)}{\tau_R s} [\theta_p(s) - \theta_m(s)] \quad (12.3)$$

Considering next the system of Figure 12.2 with feedforward compensation, let us write the equation for the summer output:

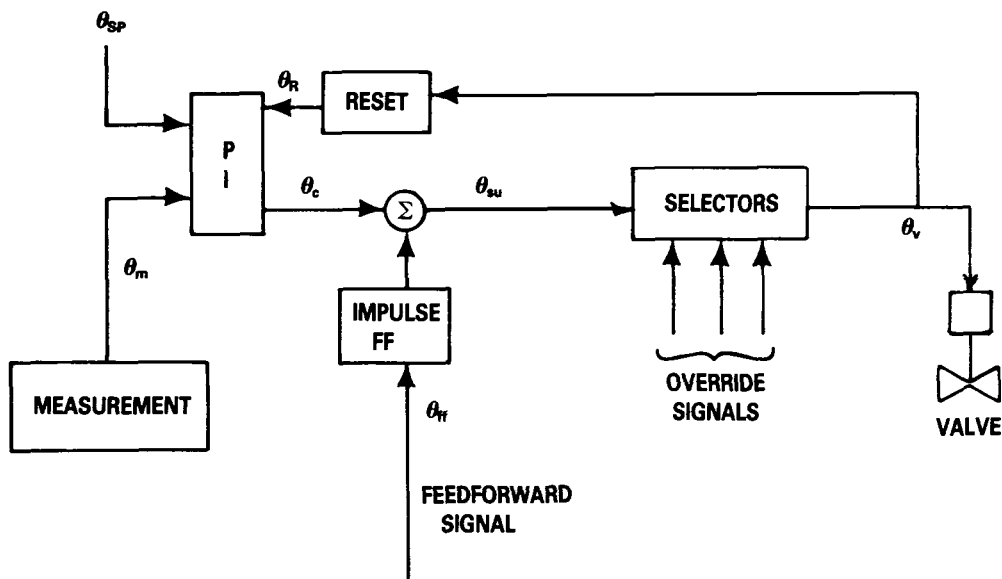


FIGURE 12.2
Single-loop with feedforward compensation

$$\theta_{su}(s) = \theta_c(s) + \theta_{ff}(s) \frac{K_{ff} \tau_{ff} s}{\tau_{ff} s + 1} \quad (12.4)$$

If we assume that there are no overrides:

$$\theta_{su}(s) = \theta_v(s)$$

Letting:

$$[\theta_{sp}(s) - \theta_m(s)] = \varepsilon(s) \quad (12.5)$$

we may combine equations (12.1), (12.2), (12.4), and (12.5) to get:

$$\theta_{su}(s) = K_c \varepsilon(s) + \frac{\theta_{su}(s)}{\tau_R s + 1} + \theta_{ff}(s) \frac{K_{ff} \tau_{ff} s}{\tau_{ff} s + 1} \quad (12.6)$$

which reduces to:

$$\theta_{su}(s) = K_c \frac{\tau_R s + 1}{\tau_R s} \varepsilon(s) + K_{ff} \frac{\tau_{ff} \tau_R s + 1}{\tau_R \tau_{ff} s + 1} \theta_{ff}(s) \quad (12.7)$$

If we make $\tau_{ff} = \tau_R$, this reduces further to:

$$\theta_{su}(s) = K_c \frac{\tau_R s + 1}{\tau_R s} \varepsilon(s) + K_{ff} \theta_{ff}(s) \quad (12.8)$$

With this design steady-state signals are not fed forward, but *changes* are. If, for example:

$$\theta_{ff}(s) = \frac{\Delta\theta_{ff}}{s}$$

which is a step function, then $\theta_{su}(s)$ also shows a step change.

Occasionally one finds a single loop with external set, that is, the set point is adjusted from another device. This usually occurs when one wishes to control a variable whose set point must be calculated. For example, one might want to maintain a constant holdup time in a chemical reactor and use throughput flow to calculate desired liquid level.

Perhaps, however, the most common example is open-loop flow-ratio control. A “wild” flow is multiplied by a constant to calculate the set point of the controller of the manipulated flow. This is not cascade control since only one controller is involved.

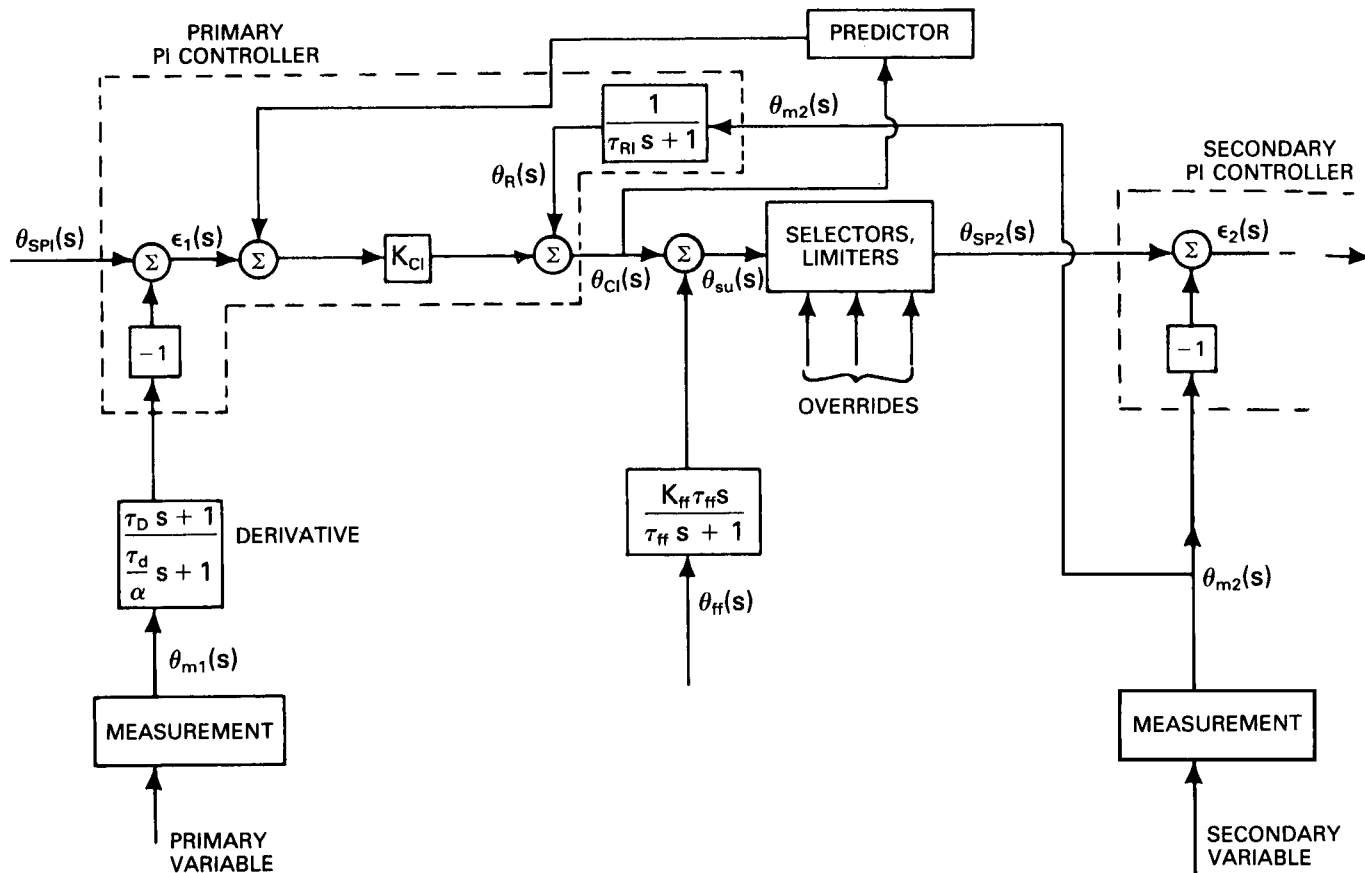
A true cascade system (see Figure 12.3) has at least two controllers: a primary or master one and a secondary or slave one, whose set point is normally set by the primary controller. An example would be column-top temperature control cascaded to reflux flow control. Our preferred scheme uses the secondary variable for external reset feedback to the primary controller. This greatly minimizes the need for manual–automatic switching—none is needed in the primary control station.

Cascade control probably has been used many times when feedforward compensation would provide superior control. For stability reasons the secondary loop in a cascade system should be five to ten times faster than the primary loop and should have dead-beat (nonresonant) tuning. But we increasingly cascade to flow controls that are usually so much faster than the primary loops that there are no stability problems.

Use of the Smith and other predictors recently has been shown by Seborg et al.³ to provide substantially better feedback control of distillation-column terminal compositions than do PI and PID controllers. Further work needs to be done to compare their performance with that achieved by PI or PID controllers with feedforward compensation.

12.4 ADJUSTMENT OF CONTROLLER PARAMETERS (CONTROLLER TUNING)

In Chapter 1 we suggested a control system structure in which most control loops are divided into three classes: (1) material balance, (2) product quality, and (3) constraints. If we look at these loops from the standpoint of classical, single-loop theory, we see that the material-balance loops function as low-pass filters (slow dynamics) while the product-quality loops function as bandpass or high-pass filters (faster dynamics).

**FIGURE 12.3****Primary controller with optional enhanced control features—cascade control system**

To eliminate interaction between the two, we design the product-quality loops to have high, closed-loop natural frequencies compared with those of the material-balance loops. This design has two advantages. First, it eliminates interactions between the two classes of loops; second, it provides great attenuation for disturbances that might otherwise upset product quality. The material-balance loops not only filter out flow disturbances, but also filter out composition disturbances if the holdups are well agitated.

For controller tuning we therefore follow two basic philosophies: (1) that for averaging controls such as most liquid-level and some pressure controls, and (2) that for all other applications.

For averaging level control we try for a damping ratio of at least unity. This is discussed in more detail in Chapter 16. For other applications we usually strive for tuning that will make the loop slightly underdamped. An M_p of 2db is the normal criterion, which is considerably more damped (less resonant) than the response obtained by Ziegler–Nichols and similar “tight” tuning procedures. The two Ziegler–Nichols procedures are empirical, the tests are difficult to run outside of a laboratory or computer, and the end result usually is a highly resonant system that is easily upset.

Although not usually pointed out in the literature, it is also true, as Gould⁸ has indicated, that the engineer should be careful not to design a feedback system whose closed-loop natural frequency is within the frequency range of disturbances. Feedforward compensation often may be used instead of feedback to get improved control without problems with either stability or resonance peaks.

A modern perspective on controller tuning has been presented elsewhere.⁹

12.5 ENHANCED CONTROL OF DISTILLATION COLUMNS VIA ON-LINE MODELS

Once a distillation column or train for a continuous chemical process has been designed and built, the three best opportunities for profitable operation via process control are usually: (1) maximum-capacity operation if the plant is production limited, (2) minimum-cost operation if the plant is market limited, and (3) increased annual availability (sometimes called utility). The last of these is greatly aided by constraint controls, commonly called overrides, to keep the plant operating safely and smoothly when it might otherwise be shut down by interlocks or operator decision. Maximum-capacity or minimum-cost operation is facilitated by the use of on-line models, usually implemented by a digital computer. Functionally these will be, in most cases, sophisticated feedforward compensators, or constraint controls. With these two limited exceptions, optimization is not considered in this book. Since the on-line models to be employed require calibration, some degree of on-line identification is highly desirable.

An interesting implication of studies to date is that for product quality

control we may need, at least more frequently than in the past, greatly increased sensitivity or resolution in control valves; in addition, temperature and flow measurements may require much increased accuracy—less than one part in 10,000.

The economic value of the enhanced controls discussed in this book varies widely from column to column, and there is no implication that the techniques presented are universally desirable.

Adaptive Tuning of Feedback Controllers

The feedback controllers we have in mind here are those for control of column terminal compositions. As shown in a recent paper,¹⁰ the static gains—changes in top and bottom compositions in response to changes in distillate or boilup or other variables—can be extremely nonlinear. Generally speaking, as top or bottom composition becomes purer, the static gains become smaller. As shown in reference 10, for one example $\partial x/\partial D$ (change in top composition per unit change in distillate) decreased 30 percent for less than 1 percent change in distillate flow. This can play havoc with controller tuning, particularly controller gain.

Reference 10 shows that many kinds of column gains may be calculated by a modified tray-to-tray method. For many columns it will be adequate to approximate the separation by the model for a binary system where relative volatility may be a nonlinear function of composition. In most cases it is probably satisfactory to run the model no more frequently than every 5 or 10 minutes. Since effective use of the model requires a knowledge of the number of theoretical trays, some sort of physical calibration or identification is required.

At present it appears that only controller gain need be adaptively tuned. The work of Wahl and Harriott¹¹ suggests that column dynamics are relatively insensitive to small changes in reflux and boilup. This means that reset and derivative (if used) need not be changed. Therefore, after the computer has calculated the appropriate column gain for a particular loop, it can easily calculate required controller gain to hold loop gain constant.

In the absence of adaptive tuning, feedback controller gain must be set low enough that the loop never becomes unstable. This means that much of the time the loop will be overdamped and sluggish. Composition, therefore, will be more variable. To ensure getting a good product, production personnel will usually adjust the composition set point for higher purity. This means an increase in average utility consumption and lower maximum column capacity.

Feedforward Compensation

Our use of feedforward compensation for distillation columns currently is based almost entirely on feed-rate changes. Thus we have reflux-to-feed, steam-to-feed, and sometimes other ratio-control systems. But there are at least three other variables that can be important: feed composition, feed enthalpy or q , and column pressure.

By use of the tray-to-tray model mentioned previously, one may calculate the required changes in distillate or reflux at column top, and boilup or bottom-product flow at column base required to hold terminal composition constant in the face of changes in the variables listed.

The three great advantages accruing from the use of feedforward compensation are:

1. Compositions are held closer to set points than is possible with unaided feedback controllers.
2. There are no problems with stability such as sometimes occur with tightly tuned, sophisticated feedback controllers, or with cascade controls.
3. Maximum column capacity is increased since manipulated variables—reflux or distillate, boilup or bottom product—are not changed as much as if unaided feedback controllers were used. Similarly, the column is less likely to weep or dump.

In work carried out at the University of Delaware a few years ago, Rippin and Lamb¹² showed that most of the benefit of feedforward is from the static gain term. Fairly simple feedforward dynamic functions such as first-order lead-lag are usually adequate and may be calculated off line.

More Accurate Determination of Constraints

The most important constraints here are those for flooding, entrainment, dumping, and maximum and minimum column-feed rates.

Ordinarily our approach to protection against flooding is to use a high column ΔP override to pinch steam. The set point is usually determined in advance by consultation with the column designer. To protect against dumping or weeping, we usually provide a minimum steam-flow override whose set point again is determined by discussion with the column designer.

Flooding and dumping, however, are really functions of a number of variables. More column capacity could be achieved by using on-line models to calculate maximum permissible ΔP or minimum required steam flow.¹³

If, for a given column, the accuracy of these models is not sufficient, we will have to obtain the data experimentally. Usually we test for flooding by fixing as many column operating variables as possible and gradually increasing boilup until a rapid increase in column ΔP or a rapid decrease in column base level occurs. We could also check for flooding on line. From material-balance and energy-balance equations, we can calculate internal reflux. Upon knowing boilup and bottom-product flow, we can determine whether base level should be increasing or decreasing and at what rate. A more rapid drop in level than predicted would indicate loss of internal reflux, and therefore flooding.

A more positive approach is to measure internal reflux, not only to the column base, but also to each feed tray, and from each tray where there is a side draw. A possible way to accomplish this is to design the downcomer bottom reservoir so that its outlet weir is a Sutro weir. The outflow from such

a weir is linearly proportional to liquid height above the weir reference line. A differential-pressure transmitter, therefore, would give an output signal linearly proportioned to flow.

If reflux flow thus calculated becomes significantly lower than predicted by the column material and energy-balance model, flooding is indicated. We can also use such flow measurements to determine column inverse response.¹⁴ Dumping is indicated if base level rises much faster than predicted by the column model.

The closer to flooding the column can operate, the greater is the column capacity. The closer to dumping the column can run, the greater is the column turndown. For best results column feed rate must be limited. As maximum permissible column ΔP is approached, the column feed rate should not exceed that which the column can handle (i.e., can separate properly without flooding). Similarly, as the column operation approaches dumping, feed rate should not fall below a minimum to avoid using excessive steam per pound of feed. Both maximum and minimum feed rate may be calculated by the column model.

Entrainment is related to flooding and there are models for its prediction. Sometimes, however, entrainment becomes dominant before flooding occurs. That is, ordinarily one would expect each increase in reflux to result in an increase in overhead purity. But sometimes before flooding occurs a point is reached at which a further increase in reflux results in a decrease in purity. With an overhead composition measurement, this relationship may be tracked and noted for control purposes. Control at the maximum reflux rate may be achieved by a peak-seeking optimizer (see Chapter 9).

This problem is sometimes approached in another way by determining column tray efficiency on line.¹⁵ Determining tray efficiency is very desirable in order to calibrate the column model used.

Balancing Energy- and Material-Handling Capacities

This section could well have another title: "How To Use Column Pressure as a Manipulated Variable." Although the practice of using automatic pressure control at a predetermined, fixed value is well established in the chemical and petroleum industries, there is seldom a good technical or economic reason for its existence. In some of the older literature, it use is urged because otherwise column temperatures might not be good measures of composition. But temperature by itself is not a good measure of composition. Increasingly, therefore, pressure either is allowed to float—that is, to find its own level—or is manipulated either to minimize steam consumption per pound of product or to maximize column capacity.¹⁶

With the advent of heat-conservation schemes, it is particularly difficult to generalize about how to manipulate pressure most successfully. Some of the considerations, however, are as follows:

1. If the condenser capacity is limiting (compared with those of the column and reboiler), raising column pressure will increase condenser capacity. But this will reduce reboiler capacity.

2. Raising column pressure may interfere with or limit the capability of the feed system to feed the column.
3. If reboiler capacity is limiting, it may be increased by lowering column pressure.
4. Lowering column pressure may interfere with pumping away or letting down column product flows.
5. For many columns, if maximum capacity is not needed, operation at the lowest possible pressures is desirable to take advantage of higher relative volatility to achieve minimum steam consumption per pound of feed.
6. Operation at lower than design pressure may cause entrainment or flooding. Fortunately we most commonly operate with low pressures at low feed rate and higher pressures at higher feed rates.

We previously suggested (see Chapter 9) that the positions of the steam and cooling-water valves give clues as to whether the reboiler or condenser is limiting. A wide-open valve indicates that the associated heat exchanger is doing its utmost. For example, if the condenser cooling-water valve is, say, 95 percent open, we might increase column pressure to prohibit the valve from opening further.

From the preceding one can see the fairly elaborate considerations regarding the role of pressure in heat transfer, flooding, entrainment, and column material balance. The use of material- and energy-balance models, together with a pressure–relative volatility–separation model, may be required.

On-Line Identification

Design equations for process equipment rarely have an accuracy of better than ± 10 percent, and often are much worse. This means that accurate determination of capacities and operating characteristics must rely on tests of the completed plant. Since heat exchangers sometimes foul, pumps and valves wear, and columns occasionally plug, it is desirable to run tests periodically. Typical plant instrumentation is seldom adequate for such testing. For example, rarely does one find a measurement of cooling-water flow rate, let alone inlet and exit cooling-water temperatures. But tests on a water-cooled condenser are useless without this information.

With sufficient measurements an on-line computer periodically can determine:

- Heat-transfer coefficients
- Column approach to flooding and entrainment
- Column approach to dumping
- Column tray efficiency
- Column inverse response
- Column-material and energy balances

Some of the preceding information can be used by the computer directly for enhanced control and some can be used to predict the need for maintenance.

Procedures for running tests on line either by statistical analysis of operating data or by injecting test signals are commonly labeled "identification" or "parameter estimation."¹⁷ Basically a computer processes input data to do one or more of the following:

1. Determine static coefficients such as tray efficiency or heat-transfer coefficients.
2. Determine numerical values of coefficients of preselected differential equations or transfer functions.
3. Find a suitable differential equation or transfer function for a process operation where a mathematical model is lacking. Years ago this was called "black box" testing.

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13 Tray Dynamics— Material Balance*

13.1 INTRODUCTION

in the past it has been common practice to treat the response of internal reflux to either external reflux or feed changes as the result of a chain of noninteracting, first-order lags. In reality, as we shall see, it is a chain of noninteracting second-order lags, although for many columns the first-order approximation is adequate. It has also been assumed that internal reflux is not affected by vapor flow changes. We now know, however, that, depending on tray design and operating conditions, an increase in vapor flow may cause internal reflux: (1) to increase temporarily, (2) to decrease temporarily, or (3) not to change at all. The first of these is commonly termed “inverse response” because it causes a momentary increase in low boilers in the column base, which is followed by a long-term decrease in low boiler concentration. This apparently was first noted by Rijnsdorp,¹ although Harriott² observed that in an AIChE report³ a bubble cap tray at low liquid rates had less holdup at higher F factors than at low F factors.

In this chapter we will derive the dynamic relationships between internal reflux and both vapor rate and external reflux (or feed) as function of tray and column design. The basic tray hydraulic equations are based on the treatment by Van Winkle.⁴ First we discuss what happens on an individual tray, and then derive an approximate model for a combination of trays. Vapor flow will be assumed to occur without lags, and heat-storage effects will be assumed to be negligible. A discussion of reboiler dynamics will be deferred to Chapter 15.

* This chapter is based on reference 5.

13.2 TRAY HYDRAULICS (See Figure 13.1)

Downcomer Liquid Level

The height of liquid level in the downcomer is the integral of the difference between the flow from tray $n + 1$, w_{n+1} , and the downcomer outflow, w_{DC} .

In Laplace transform notation:

$$H_{DC}(s) = \frac{w_{n+1}(s) - w_{DC}(s)}{\rho_{DC} A_{DC} s} \quad (13.1)$$

where

w_{DC} = flow, lbm/sec, from downcomer,
from tray $n + 1$

w_{n+1} = overflow, lbm/sec, from tray $n + 1$,
into downcomer

ρ_{DC} = liquid density in downcomer, lbm/ft³,
liquid usually slightly aerated

A_{DC} = downcomer cross-sectional area, ft²,
assumed to be uniform

H_{DC} = liquid height, feet, in downcomer

Downcomer Pressure Drop and Flow

The downcomer pressure drop is determined by the differences between the liquid heads and static heads:

$$\Delta P_{DC}(t) = \frac{g_L}{g_c} [\rho_{DC} H_{DC}(t) - \rho_{TR}(t) H'_{TR}(t)] - [P_n(t) - P_{n+1}(t)] \quad (13.2)$$

where

ΔP_{DC} = downcomer pressure drop, lbf/ft²

ρ_{TR} = density of aerated liquid on
tray n , lbm/ft³

H'_{TR} = height, feet, of aerated liquid on
tray just downstream of inlet weir

P_{n+1} = pressure, lbf/ft², above liquid on
tray $n + 1$; P_n is pressure on tray n

Note that we assume that ρ_{DC} does not vary.

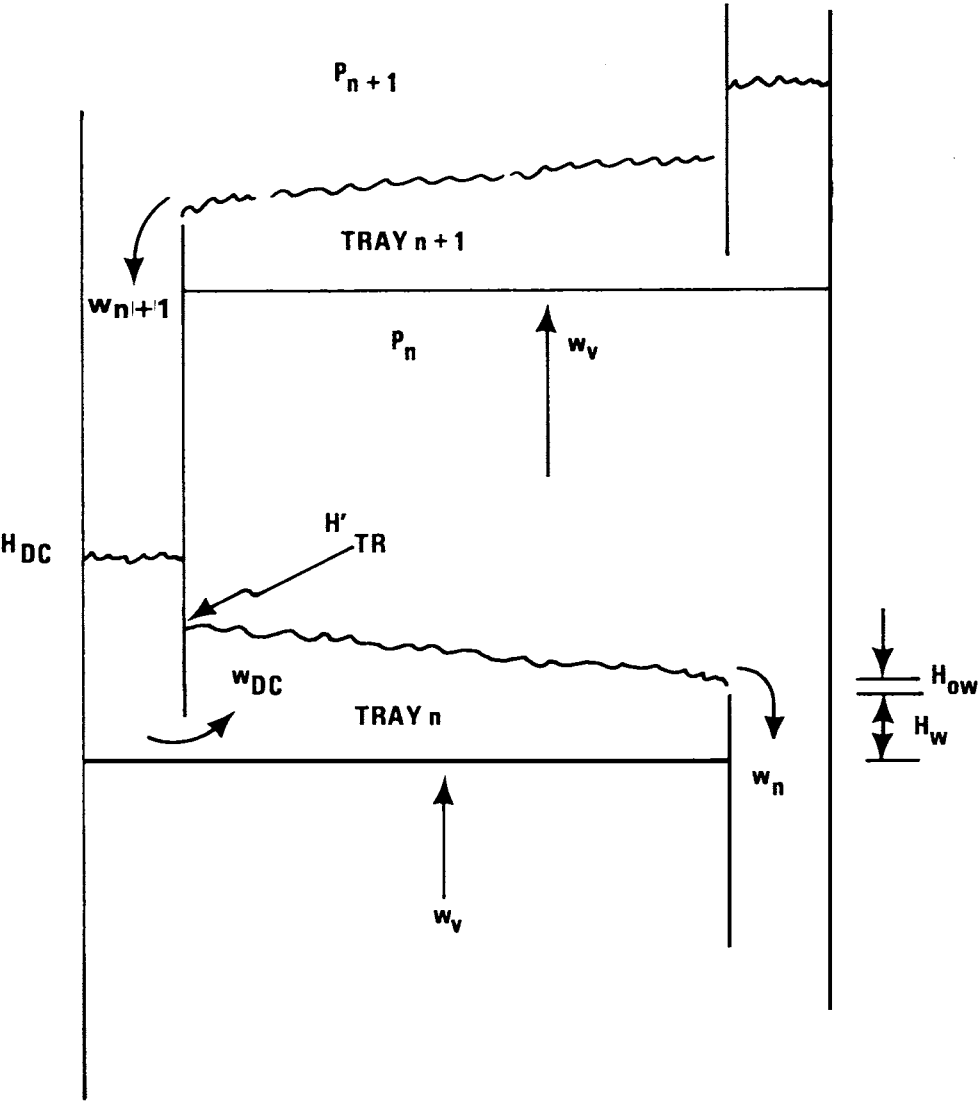


FIGURE 13.1
Distillation tray schematic for flows and liquid evaluations

Equation (13.2) may be Laplace transformed to:

$$\Delta P_{DC}(s) = \frac{g_L}{g_c} [\rho_{DC} H_{DC}(s) - \bar{\rho}_{TR} H'_{TR}(s) - \bar{H}'_{TR} \rho_{TR}(s)] - [P_n(s) - P_{n+1}(s)] \quad (13.3)$$

Next, from page 508 of Van Winkle:⁴

$$h_d = \frac{0.03 Q_{DC}^2}{(100 A_{dm})^2} \quad (13.4)$$

where

h_d = liquid head in inches of liquid-flow
pressure drop through the downcomer

Q_{DC} = downcomer flow, gpm

A_{dm} = minimum downcomer flow area, ft²

This equation may be rewritten:

$$\begin{aligned} \Delta P_{DC} &= \rho_{DC} \frac{h_d}{12} \times \frac{g_L}{g_c} \\ &= \frac{0.03}{12} \times \frac{(448.8)^2}{(100 A_{dm})^2} \times \frac{w_{DC}^2}{\rho_{DC}} \times \frac{g_L}{g_c} \end{aligned} \quad (13.5)$$

$$= \frac{0.05036 w_{DC}^2}{A_{dm}^2 \rho_{DC}} \times \frac{g_L}{g_c} \quad (13.6)$$

or

$$w_{DC} = \sqrt{\frac{\Delta P_{DC} \rho_{DC} A_{dm}^2}{0.05036}} \times \frac{g_c}{g_L} \quad (13.7)$$

$$= 4.456 A_{dm} \sqrt{\Delta P_{DC} \rho_{DC}} \frac{g_c}{g_L} \quad (13.8)$$

and

$$\frac{\partial w_{DC}}{\partial \Delta P_{DC}} = \frac{1}{2} \Delta P_{DC}^{-0.5} A_{dm} \times 4.456 \sqrt{\rho_{DC}} \frac{g_c}{g_L} \quad (13.9)$$

$$= \frac{1}{2} \frac{\bar{w}_{DC}}{\Delta P_{DC}} \quad (13.10)$$

Now since w_{DC} is determined by ΔP_{DC} , we may write:

$$w_{DC}(s) = \frac{\partial w_{DC}(s)}{\partial \Delta P_{DC}} \Delta P_{DC}(s) \quad (13.11)$$

Aerated Liquid Holdup and Gradient on Tray

With reference to Figure 13.1, we assume a linear gradient across the tray:

$$\frac{V_{TR}(t)}{A_{TR}} = \frac{[H_{ow}(t) + H_w] + H'_{TR}(t)}{2} \quad (13.12)$$

where

V_{TR} = volume, ft³, of aerated liquid
on active area of tray (downcomer
volume excluded)

A_{TR} = active tray area, ft² (area where
bubbling occurs)

H_{ow} = height, feet, over outlet weir

H_w = height, feet, of outlet weir (constant)

H'_{TR} = aerated liquid height, feet, at tray
inlet (see Figure 13.1)

$$= H_{ow} + H_w$$

or in Laplace transform notation:

$$\frac{V_{TR}(s)}{A_{TR}} = \frac{H_{ow}(s) + H'_{TR}(s)}{2} \quad (13.13)$$

Inlet Liquid Height Over Weir

Let us assume that the change in inlet height over weir is the same as the change in outlet height over weir. Then:

$$H_{ow}(s) = H'_{TR}(s) \quad (13.14)$$

Volume of Liquid on Tray

The volume of liquid on a tray is related to the inventory and the density:

$$V_{TR}(t) = \frac{W_{TR}(t)}{\rho_{TR}(t)} \quad (13.15)$$

where W_{TR} is the active tray holdup, lbm. By Laplace transforming we obtain:

$$V_{TR}(s) = \frac{\partial V_{TR}}{\partial W_{TR}} W_{TR}(s) + \frac{\partial V_{TR}}{\partial \rho_{TR}} \rho_{TR}(s) \quad (13.16)$$

where

$$\frac{\partial V_{TR}}{\partial W_{TR}} = \frac{1}{\bar{\rho}_{TR}} \quad [\text{From equation (13.15)}]$$

and

$$\frac{\partial V_{TR}}{\partial \rho_{TR}} = \frac{-\bar{W}_{TR}}{\bar{\rho}^2} = \frac{-\bar{V}_{TR}}{\bar{\rho}_{TR}} \quad [\text{From equation (13.15)}]$$

Tray Liquid Material Balance

On the active part of the tray, the liquid material balance in Laplace transform notation is:

$$\frac{w_{DC}(s) - w_n(s)}{s} = W_{TR}(s) \quad (13.17)$$

where w_n = overflow, lbm/sec, from tray n .

Aerated Liquid Density as a Function of Vapor Velocity

In Figure 13.16, page 516, Van Winkle⁴ presents a plot that shows that froth density decreases with an increase in vapor rate. From this we can calculate a slope, $\partial \rho_{TR} / \partial F$. Then:

$$\rho_{TR}(s) = \frac{\partial \rho_{TR}}{\partial F} \frac{\partial F}{\partial w_v} w_v(s) \quad (13.18)$$

where

$$F = F \text{ factor}$$

$$= U_{VA} \rho_v^{0.5} = \frac{w_v}{A_{TR}} \rho_v^{-0.5}$$

$$U_{VA} = \text{vapor velocity, ft/sec, corresponding to } A_{TR}$$

$$w_v = \text{vapor rate, lbm/sec}$$

$$\rho_v = \text{vapor density, } \frac{\text{lbm}}{\text{ft}^3}$$

$$\frac{\partial F}{\partial w_v} = \frac{\bar{\rho}_v^{-0.5}}{A_{TR}}$$

Liquid Overflow from Tray

The Francis weir formula can be written in the form:

$$Q(t) = \frac{w_n(t)}{\rho_{TR}(t)} = k H_{ow}^{3/2} \quad (13.19)$$

or

$$w_n(t) = \rho_{TR}(t) k H_{ow}^{3/2} \quad (13.20)$$

where

Q = overflow, ft³/sec, of aerated liquid

k = a constant for any given weir (see
Van Winkle,⁴ pages 507–508)

In Laplace transform notation:

$$w_n(s) = \frac{\partial w_n}{\partial \rho_{TR}} \rho_{TR}(s) + \frac{\partial w_n}{\partial H_{ow}} H_{ow}(s) \quad (13.21)$$

where

$$\frac{\partial w_n}{\partial \rho_{TR}} = k(\bar{H}_{ow})^{3/2} = \frac{\bar{w}_n}{\rho_{TR}} \quad (13.22)$$

and

$$\begin{aligned} \frac{\partial w_n}{\partial H_{ow}} &= k \bar{\rho}_{TR} \times \frac{3}{2} (\bar{H}_{ow})^{0.5} \\ &= \frac{3}{2} \frac{\bar{w}_n}{2 \bar{H}_{ow}} \end{aligned} \quad (13.23)$$

Tray Pressure Drop

In Laplace transform notation:

$$P_n(s) - P_{n+1}(s) = \Delta P_{TR}(s) = \frac{\partial \Delta P_{TR}}{\partial w_v} w_v(s) \quad (13.24)$$

For perforated trays Van Winkle (page 519 of reference 4) gives the relationship:

$$\Delta P_{TR} = \frac{\rho_L h_o}{12} \left(\frac{g_L}{g_c} \right) = \frac{0.186 U_h^2}{12 C_o^2} \rho_v \left(\frac{g_L}{g_c} \right) \quad (13.25)$$

where

U_h = velocity, ft/sec, through the holes

$$= \frac{w_v}{\rho_v A_h}$$

A_h = total hole area per tray, ft²

h_o = dry-plate pressure drop, inches of
clear liquid

ρ_L = density, lbm/ft³, of clear liquid

ρ_v = vapor density, lbm/ft³

C_o = discharge coefficient that is a function of both the tray-thickness/hole-diameter ratio, and the hole-area/active-area ratio. This is presented in Figure 13.18, page 519, of Van Winkle.⁴

Thistlethwaite⁶ has developed a correlation for C_o that facilitates computations:

$$C_o = 0.92101 - 0.42305 \frac{T_t}{D_h} + 0.11055 \ln\left(\frac{T_t}{D_h}\right) + 0.2935 \left(\frac{T_t}{D_h}\right)^2 + 0.72 \left(\frac{A_h}{A_{TR}} - 0.5\right)$$

where T_t = tray thickness and D_h = hole diameter.

From the above:

$$\Delta P_{TR} = 0.0155 \left(\frac{w_v}{A_h}\right)^2 \frac{1}{\rho_v C_o^2} \left(\frac{g_L}{g_c}\right) \quad (13.26)$$

so

$$\frac{\partial \Delta P_{TR}}{\partial w_v} = \frac{0.0310 \bar{w}_v}{A_h^2 \rho_v C_o^2} \quad (13.27)$$

$$= 2 \left(\frac{\Delta P_{TR}}{\bar{w}_v} \right) \quad (13.28)$$

For valve trays, if the caps are not fully lifted:

$$\frac{\partial \Delta P_{TR}}{\partial w_v} \simeq 0$$

Once the caps have been fully lifted, the equation for perforated trays applies.

13.3 DERIVATION OF OVERALL TRAY EQUATION

As a first step in deriving an overall tray equation, let us construct a signal flow diagram from equations (13.1), (13.3), (13.11), (13.13), (13.14), (13.16), (13.17), (13.18), (13.21), and (13.24). This is shown in Figure 13.2. By successive reductions this signal flow diagram can be reduced until the following equation is derived:

$$w_n(s) = \frac{1}{\tau_{Q^2}^2 s^2 + 2\zeta \tau_{Q^2} s + 1} \left[w_{n+1}(s) + s w_v(s) \right. \\ \left. \times \left(\left[-\bar{V}_{TR} \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} + \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} \frac{2}{3} \bar{H}_{ow} A_{TR} \right] \tau_{DC} s \right) \right] \quad (13.29)$$

$$+ \left[- \frac{\partial \Delta P_{TR}}{\partial w_v} - \frac{g_L}{g_c} \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} \bar{H}'_{TR} \right] \frac{g_c}{g_L} A_{DC} \\ + \left[- \bar{V}_{TR} \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} + \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} \frac{2}{3} \bar{H}_{ow} (A_{TR} + A_{DC}) \right] \Bigg] \Bigg]$$

where

$$\tau_{DC} = \frac{\frac{g_c}{g_L} A_{DC}}{\frac{\partial w_{DC}}{\partial \Delta P_{DC}}} \\ \tau_Q = \sqrt{\frac{\frac{\rho_{TR} A_{TR}}{3} \frac{\bar{w}_n}{\bar{H}_{ow}}}} \\ \zeta = \frac{1}{2\tau_Q} \left(\frac{\rho_{TR} (A_{TR} + A_{DC}) + \frac{3}{2} \frac{\bar{w}_n}{\bar{H}_{ow}} \tau_{DC}}{\frac{3}{2} \frac{\bar{w}_n}{\bar{H}_{ow}}} \right)$$

For all columns examined so far, the denominator quadratic has factored into two terms with substantially unequal time constants. The smaller, in most cases, has been roughly equal to the numerator time constant in the multiplier for $w_v(s)$. Then equation (13.29) reduces to:

$$w_n(s) = \frac{1}{\tau_{TR}s + 1} w_{n+1}(s) + \frac{K_{TR}s}{\tau_{TR}s + 1} w_v(s) \quad (13.30)$$

where

$$\tau_{TR} \approx \frac{\bar{\rho}_{TR} (A_{TR} + A_{DC}) + \frac{3}{2} \frac{\bar{w}_n}{\bar{H}_{ow}} \tau_{DC}}{\frac{3}{2} \frac{\bar{w}_n}{\bar{H}_{ow}}} \quad \text{for } \zeta \gg 1 \quad (13.31)$$

$$K_{TR} = \left[- \frac{\partial \Delta P_{TR}}{\partial w_v} - \frac{g_L}{g_c} \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} \bar{H}'_{TR} \right] \frac{g_c}{g_L} A_{DC} \\ + \left[- \bar{V}_{TR} \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} + \frac{\partial F}{\partial w_v} \frac{\partial \rho_{TR}}{\partial F} \frac{2}{3} \bar{H}_{ow} (A_{TR} + A_{DC}) \right] \quad (13.32)$$

Note that K_{TR} has the dimensions

$$\frac{\text{lbm/sec}}{\text{lbm/sec}} \times \text{sec}$$

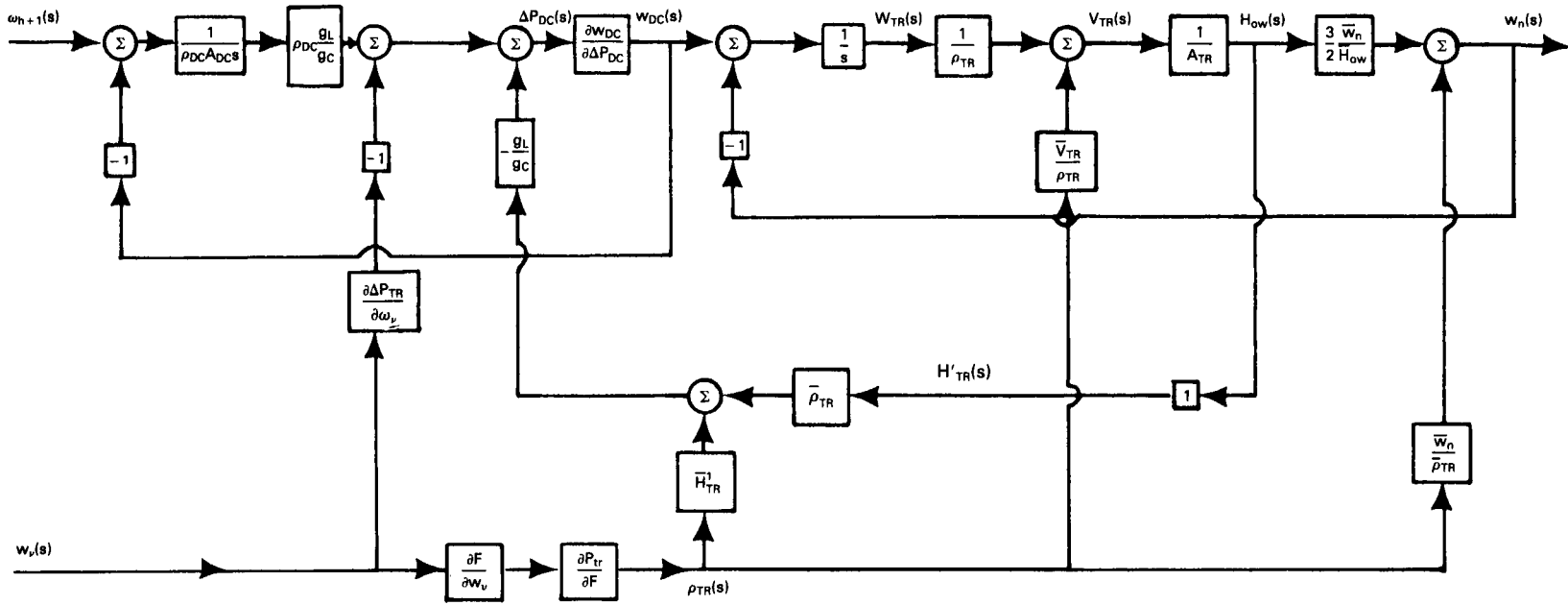


FIGURE 13.2
Preliminary signal flow diagram for tray material balance dynamics

We can now make three generalizations about K_{TR} and w_v :

1. If K_{TR} is *positive*, an *increase* in w_v causes a temporary increase in internal reflux. This results in an "inverse" response, since low boiler concentration is increased temporarily in the base of the column.
2. If K_{TR} is zero, a change in w_v causes no change in internal reflux. This we term "neutral" response.
3. If K_{TR} is *negative*, an *increase* in w_v causes a temporary *decrease* in internal reflux. This we have termed "direct" response; it temporarily augments the long-term decrease in base low boilers caused by increasing w_v .

The term that seems to affect the sign of K_{TR} most strongly is $\partial \Delta P_{TR} / \partial w_v$. This term is almost zero for valve trays, and for all valve-tray columns we have checked so far, the calculations predict inverse response over the entire range of normal operation. For sieve trays, $\partial \Delta P_{TR} / \partial w_v$ is small at low boilup rates and the calculations predict inverse response. This term increases rapidly with boilup, however [see equation (13.27)] and the calculations indicate that as w_v increases, the column shows next neutral response, and finally, for large values of w_v , direct response.

13.4 MATHEMATICAL MODEL FOR COMBINED TRAYS

A column with a number of trays may be represented by a mathematical model such as shown in Figure 13.3. This may be simulated readily on a large digital computer. For a mathematical analysis, however, or for hand calculations, we need a simpler model.

A system with a large number of identical first-order lags may be represented by a dead time a , which is equal to $n\tau$ where n is the number of lags. Therefore, the response of reflux from the lowest tray, w_1 , to a change in external reflux w_R n trays away is:

$$\frac{w_1(s)}{w_R(s)} \simeq K_{sc} e^{-as} \quad (13.33)$$

where

$$K_{sc} = 1 + \frac{c_p}{\lambda} (T_o - T_R)$$

T_o = condensing temperature, °C

T_R = external reflux temperature, °C

w_R = external reflux, lbm/sec

A simplified model for response to w_v may be arrived at by a somewhat intuitive method. Consider the overflow response of an individual tray to a

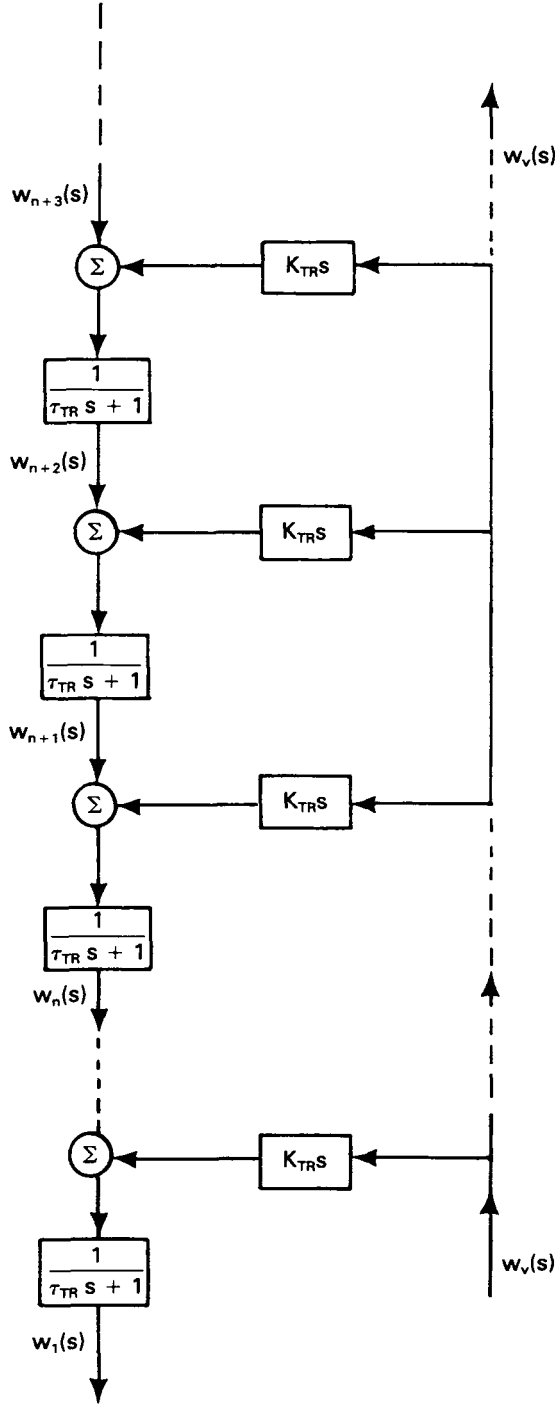


FIGURE 13.3
Material balance coupling with vapor and liquid flow

step change in w_p :

$$w(t) = \Delta w_p \frac{K_{TR}}{\tau_{TR}} e^{-t/\tau_{TR}} \quad \text{lbm/sec} \quad (13.34)$$

The integrated outflow is:

$$W(t) = \Delta w_p K_{TR} (1 - e^{-t/\tau_{TR}}) \quad (13.35)$$

At $t = \infty$:

$$W(\infty) = \Delta w_p K_{TR} \quad (13.36)$$

For n trays the total amount of liquid that is either displaced or held up is:

$$\Sigma W(\infty) = n \Delta w_p K_{TR} \quad (13.37)$$

Since we are assuming that Δw_p is felt by all of the trays simultaneously, equations (13.29) and (13.30) apply to all trays at the same time. But the outflow from each tray must flow through all lower trays. The time required for the bulk of the liquid displaced or held up to flow down is approximately $n\tau_{TR}$. Therefore, the average outflow rate is:

$$w_{\text{avg}} = \frac{n \Delta w_p K_{TR}}{n \tau_{TR}} \quad (13.38)$$

Overall, then, we obtain the approximate transfer function for an n -tray column:

$$\frac{w_1(s)}{w_p(s)} \simeq \frac{K_{TR}}{\tau_{TR}} [1 - e^{-n\tau_{TR}s}] \quad (13.39)$$

It was previously indicated that if the sign of K_{TR} is positive, the column-base composition will exhibit an inverse response to a change in boilup. Equation (13.39) indicates the possibility of another kind of inverse response, that of column-base level. If K_{TR} is positive and if K_{TR}/τ_{TR} is greater than unity, an *increase* in boilup will result in a temporary *increase* in base level. As will be discussed in Chapter 16, this can cause great difficulties where base level is controlled by boilup.

If K_{TR}/τ_{TR} is positive and close to unity, a change in boilup causes no change in level for a period of time equal to $n\tau_{TR}$. The control system seems afflicted with dead time, but in reality, as shown by equation (13.39), it is not. Thistlethwaite⁶ has carried out a more extensive analysis of inverse response in distillation columns.

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14

Distillation- Column Material-Balance Control

14.1 MATHEMATICAL MODEL—OPEN LOOP

in this chapter we will look at the relationship between level control in the overhead condensate receiver and that in the column base for several different column-control schemes. Since flows are commonly measured in pounds/unit time, we will use these units instead of molar ones. Later, in Chapter 16, we will look at individual level controls in more detail.

To illustrate mathematical modeling for column material-balance control, let us first use the conventional column of Figure 14.1. The feed, w_F , is split by the column into two parts: top product, w_D , and bottom product, w_B . It is assumed that vent losses overhead are negligible. It is further assumed that the heat-transfer dynamics of both the condenser and the reboiler are negligible; this will be true for most columns. Let us start at the column base and work up. For convenience the equations are written in Laplace transform notation.

Column Base, Including Reboiler

$$\frac{w_1(s) - w_r(s) - w_B(s)}{s} = W_B(s) \quad (14.1)$$

where

w_1 = liquid flow from first tray, lbm/min

w_r = vapor flow leaving column base, lbm/min

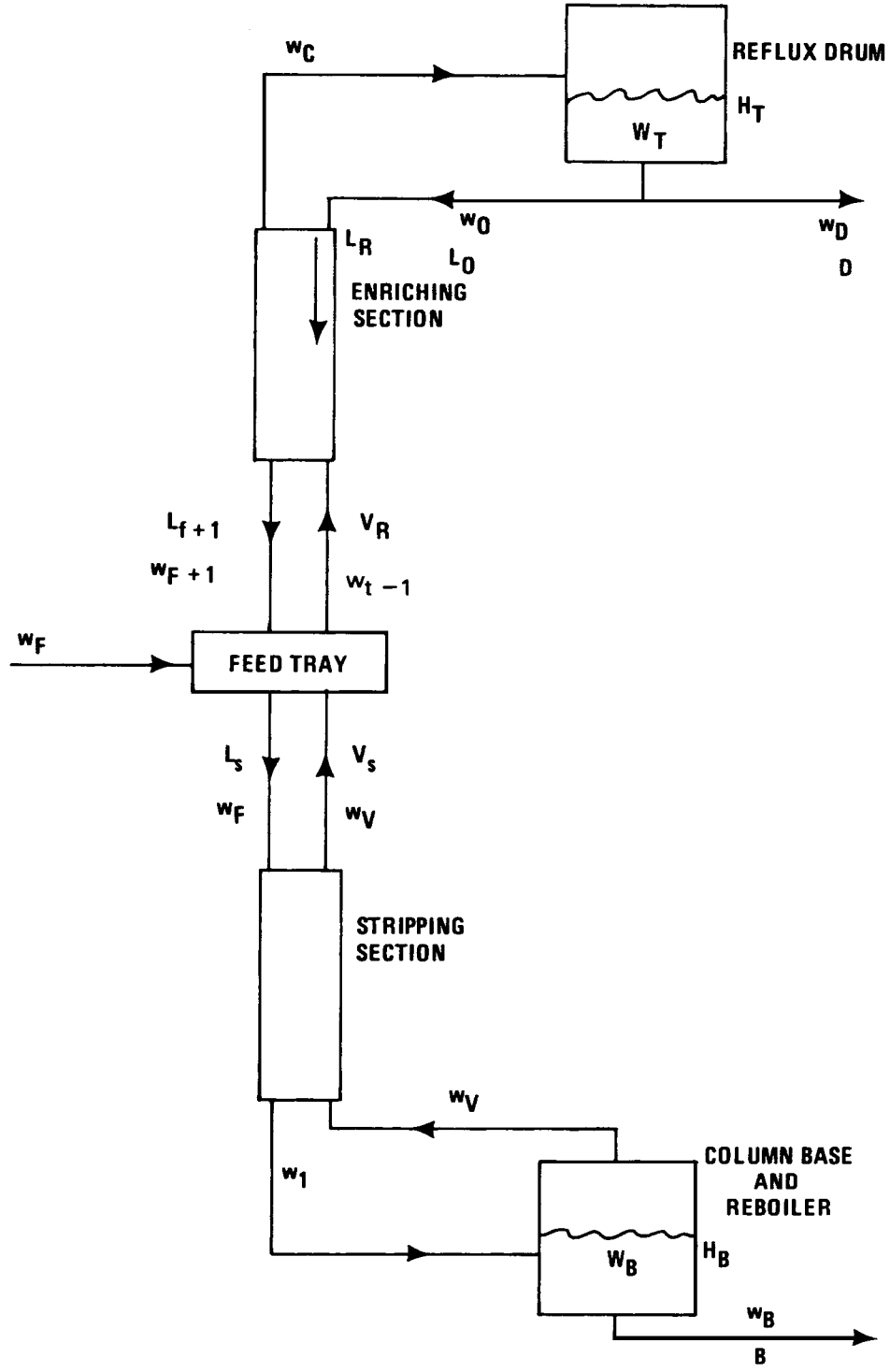


FIGURE 14.1
Distillation column material balance

w_B = bottom product flow, lbm/min

W_B = liquid inventory in column base, lbm,
within the level transmitter span, $(\Delta H_T)_B$

s = Laplace transform variable

Next:

$$\frac{W_B(s)}{\rho_B A_B} = H_B(s) \quad (14.2)$$

where

ρ_B = density, lbm/ft³ of liquid in column base

A_B = cross-sectional area of column base, ft²

H_B = liquid level in base, feet

If the reboiler is heated by steam:

$$w_v(s) = \frac{\lambda_{ST}}{\lambda_{PB}} w_s(s) \quad (14.3)$$

where

λ_{ST} = latent heat of steam, pcu/lbm

λ_{PB} = latent heat of process fluid,
pcu/lbm, at column base

w_s = steam flow, lbm/min

Usually vapor flow changes are propagated up the column very rapidly. Therefore, no great error is introduced by assuming that they appear instantaneously at the column top.

Feed Tray

The feed-tray material balance is usually written in terms of molar flows:

$$V_R = F(1 - q) + V_s \quad (14.4)$$

where

V_R = vapor flow leaving feed tray, mol/min

F = feed flow, mol/min

V_s = vapor flow entering feed tray, mol/min

q = enthalpy factor

= $\frac{(\text{Enthalpy of feed as vapor at dew point} - \text{Actual feed enthalpy})}{\text{Molar latent heat of vaporization of feed}}$

= $\frac{(L_S - L_R)}{F}$

L_R = liquid flow from top tray, mol/min

L_S = liquid flow from feed tray, mol/min

Note that if the feed is a liquid at its boiling point, $q = 1$ and $V_R = V_S$. If the feed is subcooled liquid, q is greater than 1 and V_R is less than V_S .

It is also true that:

$$V_S + F + L_{f+1} = L_f + V_R \quad (14.5)$$

From the definition of q above we can write:

$$L_{f+1} - L_f = -Fq \quad (14.6)$$

In terms of weight units equation (14.6) becomes:

$$\frac{w_{f+1}}{w_{f+1}/L_{f+1}} - \frac{w_f}{w_f/L_f} = \frac{-w_F}{w_F/F} q \quad (14.7)$$

Note that:

$$w_F = \text{feed, lbm/min}$$

$$w_f = \text{liquid flow from the feed tray, lbm/min}$$

From equation (14.7):

$$w_f = \left[\frac{w_{f+1}}{w_{f+1}/L_{f+1}} + \frac{w_F}{w_F/F} q \right] \left(\frac{w_f}{L_f} \right) \quad (14.8)$$

Since (w_{f+1}/L_{f+1}) , (w_F/F) , and (w_f/L_f) are constants, we can rewrite equation (14.8) in Laplace transform notation:

$$\begin{aligned} w_f(s) &= \left(\frac{w_f}{L_f} \right) \left(\frac{L_{f+1}}{w_{f+1}} \right) w_{f+1}(s) + \left(\frac{w_f}{L_f} \right) \left(\frac{F}{w_F} \right) \bar{q} w_F(s) \\ &\quad + \left(\frac{w_f}{L_f} \right) \left(\frac{F}{w_F} \right) \bar{w}_F q(s) \end{aligned} \quad (14.9)$$

or

$$w_f(s) = k_2 w_{f+1}(s) + k_3 w_F(s) + k_4 q(s) \quad (14.10)$$

Usually $k_2 = 1$, $k_3 = q$, and $k_4 = w_F$, although the last may not be true if the feed composition is radically different from feed-tray composition.

In going back to equation (14.4) and expressing it in weight units, we obtain:

$$\frac{w_{t-1}}{w_{t-1}/V_R} = \frac{w_F}{w_F/F} (1 - q) + \frac{w_p}{w_p/V_S} \quad (14.11)$$

whence

$$\begin{aligned} w_{t-1}(s) &= \left(\frac{w_{t-1}}{V_R} \right) \left(\frac{V_S}{w_p} \right) w_p(s) + (1 - \bar{q}) \left(\frac{F}{w_F} \right) \left(\frac{w_{t-1}}{V_R} \right) w_F(s) \\ &\quad - \left(\frac{w_{t-1}}{V_R} \right) \bar{w}_F \left(\frac{F}{w_F} \right) q(s) \end{aligned} \quad (14.12)$$

or

$$w_{i-1}(s) = k_6 w_v(s) + k_7 w_F(s) - k_8 q(s) \quad (14.13)$$

Usually $k_6 = 1$, $k_7 = 1 - q$, and $k_8 = w_F$, although the last two may not hold if feed composition is radically different from feed-tray composition.

Stripping-Section Liquid-Flow Dynamics

The transfer function between w_1 and w_f is:

$$\frac{w_1(s)}{w_f(s)} = G_2(s) \quad (14.14)$$

where $G_2(s)$ is the cumulative effect of the individual tray hydraulic lags, each with a hydraulic lag, τ_{TR} (no inverse response assumed):

$$G_2(s) = \frac{1}{(\tau_{TR}s + 1)^{N_S}} \approx e^{-a_2 s} \quad (14.15)$$

where

N_S = number of trays from the column
base to the feed tray

$$a_2 = N_S \tau_{TR}$$

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$$\frac{w_{f+1}(s)}{w_{1R}(s)} = G_1(s) \quad (14.16)$$

where

w_{1R} = liquid flow (internal reflux)
from top tray, lbm/min

$$G_1(s) = \frac{1}{(\tau_{TR}s + 1)^{N_R}} \approx e^{-a_1 s} \quad (14.17)$$

where

N_R = number of trays above the feed tray

$$a_1 = N_R \tau_{TR}$$

Note that:

$$w_{1R} = K_{sc} w_R \quad (14.18)$$

where

$$K_{sc} = 1 + \frac{c_p}{\lambda_{PT}}(\bar{T}_o - \bar{T}_R)$$

w_R = external reflux flow, lbm/min

Overhead Material Balance

The vapor flow to the condenser is:

$$w_c(s) = w_{t-1}(s) - w_R(s) \left[\frac{c_p}{\lambda_{PT}}(\bar{T}_o - \bar{T}_R) \right] \quad (14.19)$$

where

λ_{PT} = latent heat of vaporization of process
fluid specific heat, pcu/lb

c_p = process fluid specific heat, pcu/lbm°C

\bar{T}_o = average vapor temperature, °C

\bar{T}_R = average external reflux temperature, °C

w_c = vapor flow to condenser, lbm/min

Note that if there is no subcooling, $w_c = w_{t-1}$.

Condensate Receiver Material Balance

$$\frac{w_c(s) - w_D(s) - w_R(s)}{s} = W_T(s) \quad (14.20)$$

where

w_D = top-product flow rate, lbm/min

W_T = condensate receiver inventory, lbm

If the condensate receiver is a vertical, cylindrical vessel:

$$\frac{W_T(s)}{\rho_T A_T} = H_T(s) \quad (14.21)$$

where

H_T = height of liquid, feet, in receiver

ρ_T = density of top product, lbm/ft³

A_T = cross-sectional area of receiver, ft²

The preceding equations now can be combined into the signal flow diagram of Figure 14.2. As can be seen, the feed flow, external reflux, steam flow, and top- and bottom-product flows are all inputs. By providing the proper additional connections, we can design any desired type of material-balance control.

Limitations of Preceding Analysis

There are three factors that limit the accuracy of the preceding analysis. The first of these relates to the phenomenon of inverse response discussed in Chapter 13. It is characteristic of valve tray columns and some sieve tray columns operating at low boilup rates. It exercises its most serious effect in those columns where base level is controlled via steam flow. If the level becomes too high, the level controller increases the steam flow. But this causes a momentary increase in base level due to the extra liquid coming down the column (also due to thermosyphon reboiler "swell"). Without proper design the level controller can become very confused. This is discussed in detail in Chapter 16.

The second limiting factor is entrainment. Normally we assume that the only way we get liquid overhead is by condensing vapor. But at high boilup rates, entrainment may be severe enough to invalidate the simple material-balance model we have developed.

The third factor is the simplified, steady-state treatment of the feed tray. For purposes of this chapter, we do not believe this introduces a serious error. Feed tray dynamics will be dealt with more rigorously in Chapter 18.

14.2 CONTROL IN THE DIRECTION OF FLOW

Let us look at a material-balance control scheme that is in the direction of flow. Let feed rate be set by averaging level control of the feed tank, let condensate receiver level set top-product flow, and let column base level set bottom-product flow. We will assume that each level controller is cascaded to the appropriate flow controller.

Overhead Level Control

The necessary additional equation (no subcooling) is:

$$w_D(s) = H_T(s)K_{mbr}K_{chr}G_{chr}(s) \times \frac{1}{K_{mfD}} \quad (14.22)$$

where

$$\begin{aligned} K_{mfD} &= \text{distillate flow-meter gain, } \frac{\text{psi}}{\text{lbm/min}} \\ &= \frac{12}{(W_D)_{\max}} \quad \text{Note: linear flow meter} \end{aligned}$$

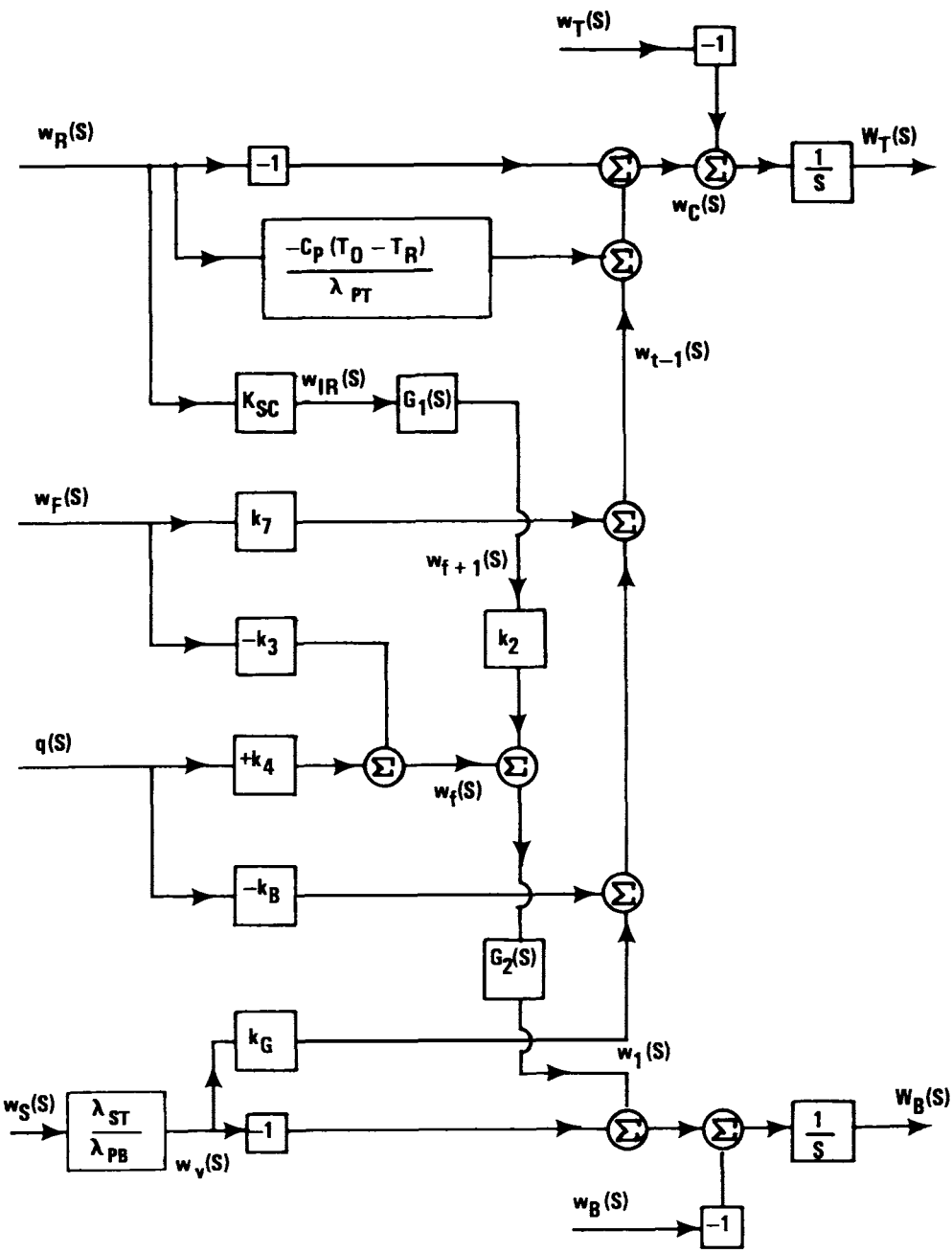


FIGURE 14.2
Signal flow diagram for column material balance

$(w_D)_{\max}$ = top-product flow-meter span, lbm/min

K_{cht} = controller gain, psi/psi

$G_{cht}(s)$ = controller dynamic gain

K_{mhT} = receiver level transmitter gain, psi/ft

Note that we have ignored the dynamics of level measurement and of the flow control loop. For averaging level control, this introduces little error.

We can now prepare the partial signal flow diagram of Figure 14.3. From this we can see by inspection that:

$$\frac{w_D(s)}{w_c(s) - w_R(s)} = \frac{1}{1 + \frac{K_{cht}G_{cht}(s)K_{mhT}}{A_T\rho_T K_{mfD}}} \quad (14.23)$$

$$= K_{HT}G_{HT}(s) \quad (14.23a)$$

For pneumatic instruments:

$$K_{mhT} = \frac{12 \text{ psi}}{(\Delta H_T)_T} \quad (14.24)$$

where $(\Delta H_T)_T$ is the level transmitter span, in feet, for a 3–15-psig output.

We can now define a characteristic time constant:

$$[\tau_H]_T = \frac{\rho_T A_T (\Delta H_T)_T}{K_{cht}(w_D)_{\max}} \quad (14.25)$$

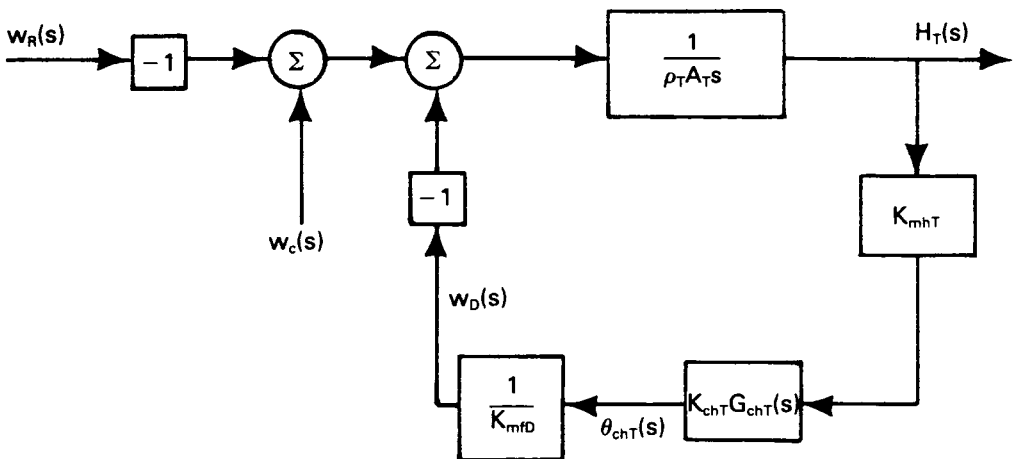


FIGURE 14.3
Signal flow diagram—condensate receiver

If a proportional-only control system is used, equation (14.23) becomes:

$$\frac{w_D(s)}{w_c(s) - w_R(s)} = \frac{1}{[\tau_H]_T s + 1} \quad (14.26)$$

while for proportional-reset level control it becomes, as indicated in Chapter 16:

$$\frac{w_D(s)}{w_c(s) - w_R(s)} = \frac{\tau_R s + 1}{\tau_R [\tau_H]_T s^2 + \tau_R s + 1} \quad (14.27)$$

where τ_R is the level controller reset time in minutes.

Usually it is desirable to have $[\tau_H]_T$ as small as convenient, say 2–5 minutes (120–300 seconds), for best control of the associated column. This is large enough to ensure that instrument and pneumatic transmission-line dynamics will not be significant. If we fix K_{ch} , then we must achieve the desired $[\tau_H]_T$ by proper choice of A_T or $(\Delta H_T)_T$ or both. Note that for best flow smoothing to another process step, one may need a larger $(\tau_H)_T$, or even an additional buffer or surge tank.

If a proportional-only controller is used, it is recommended that $K_{chT} = 2$ be chosen.

Base Level Control

The necessary additional equation is:

$$w_B(s) = \frac{1}{K_{mfB}} K_{chB} G_{chB}(s) K_{mhB} H_B(s) \quad (14.28)$$

where

$$\begin{aligned} K_{mfB} &= \text{bottom-product flow transmitter gain, } \frac{\text{psi}}{\text{lbm/min}} \\ &= 12/(\bar{w}_B)_{\max}; (\bar{w}_B)_{\max} = \text{flow-meter span, lbm/min} \\ K_{chB} &= \text{controller gain, psi/psi} \\ K_{mhB} &= \text{base-level transmitter gain, psi/ft} \end{aligned}$$

We can now prepare the partial signal flow diagram of Figure 14.4, from which we can see that:

$$\frac{w_B(s)}{w_1(s) - w_p(s)} = \frac{1}{1 + \frac{K_{chB} G_{chB}(s) K_{mhB}}{\rho_B A_B K_{mfB}}} \quad (14.29)$$

$$= K_{HFB} G_{HFB}(s) \quad (14.29a)$$

By analogy with equation (14.25) we may define:

$$[\tau_H]_B = \frac{\rho_B A_B (\Delta H_T)_B}{K_{chB} (\bar{w}_B)_{\max}} \quad (14.30)$$

where $(\Delta H_T)_B$ is the base-level transmitter span in feet for 3–15 psig output.

For proportional-only control, equation (14.29) becomes:

$$\frac{w_B(s)}{w_1(s) - w_v(s)} = \frac{1}{[\tau_H]_B s + 1} \quad (14.31)$$

while for proportional-reset control it is:

$$\frac{w_B(s)}{w_1(s) - w_v(s)} = \frac{\tau_R s + 1}{\tau_R [\tau_H]_B s^2 + \tau_R s + 1} \quad (14.32)$$

where τ_R is the level controller reset time in minutes.

It is usually desirable for best control of the associated column to make $(\tau_H)_B = 10$ –15 minutes, and if K_{chB} is specified, then the proper time constant is achieved by choice of A_B or $(\Delta H_T)_B$, or both. For proportional-only control, $K_{chB} = 2$ is recommended. Note that for best flow smoothing to another process step, one may need a large $(\tau_H)_B$, or even an additional surge or buffer tank.

We can now prepare the overall closed-loop material-balance diagram of Figure 14.5. Note that the two level controls are independent and noninteracting. If we were to add reflux/feed and steam/feed ratio controls, this statement would still be true.

14.3 CONTROL IN DIRECTION OPPOSITE TO FLOW

As an example let us choose the case of Figure 6.5 where top-product flow is the demand flow, condensate receiver level sets steam flow, base level sets feed flow, and both reflux and bottom-product flow are ratioed to top-product flow. As before, level controllers are cascaded to flow controls with linear flow meters.

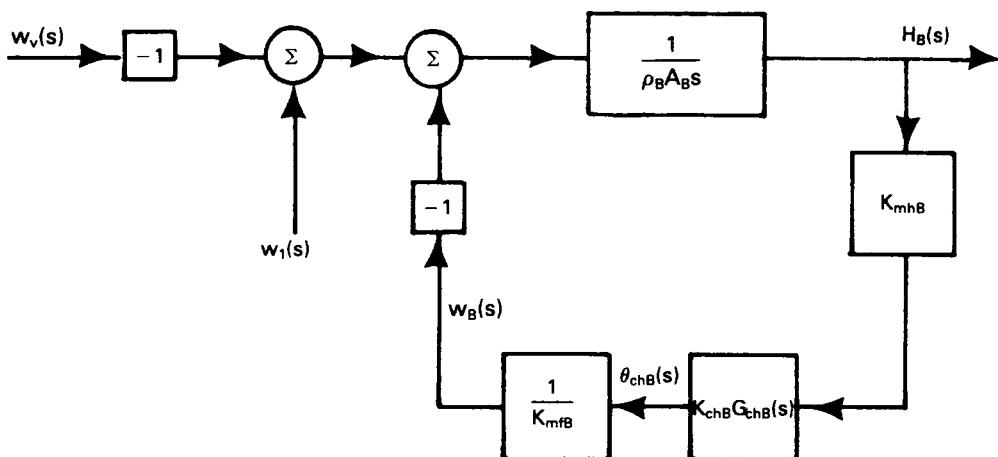


FIGURE 14.4
Signal flow diagram—column base

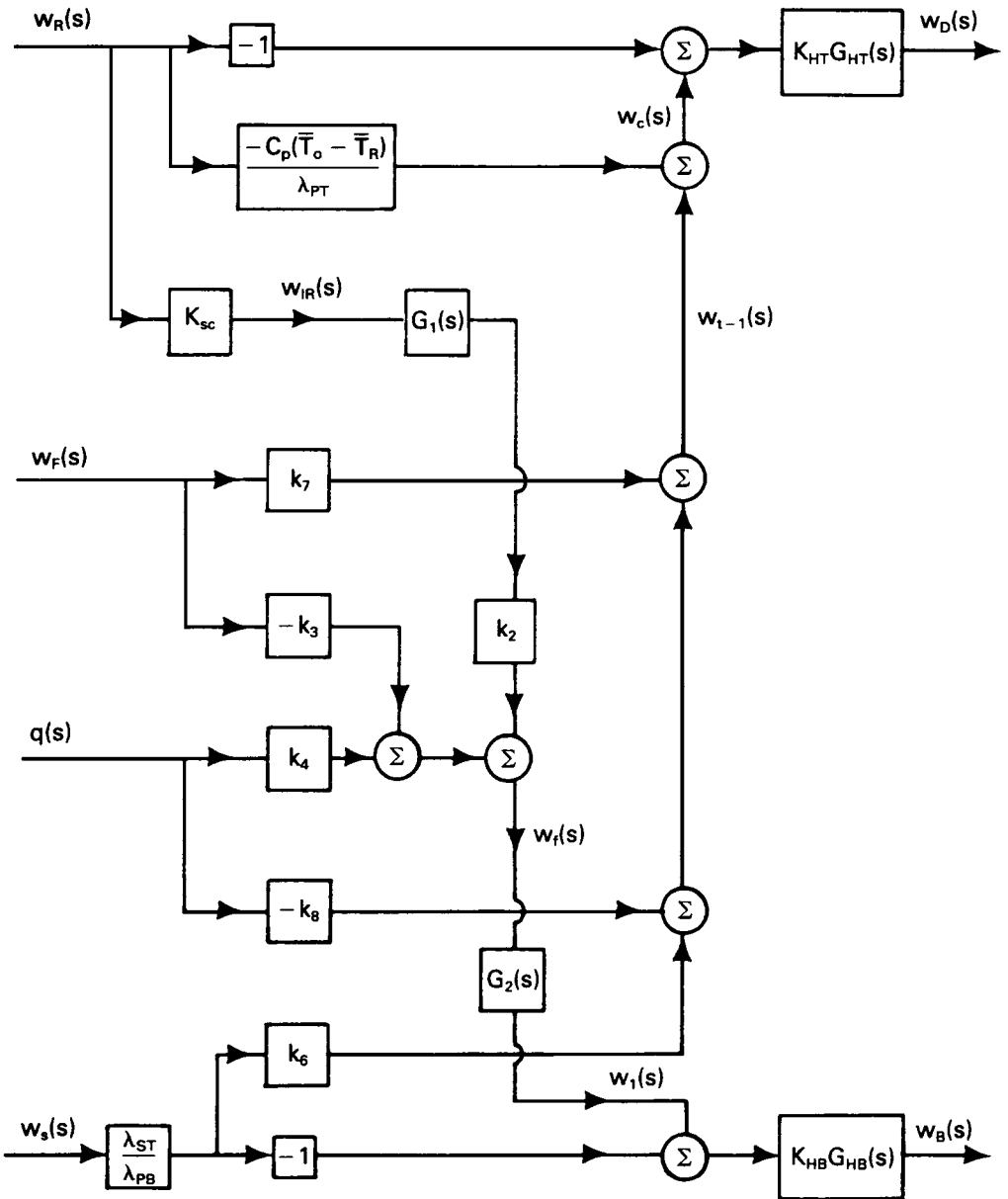


FIGURE 14.5
Signal flow diagram—material balance control in direction of flow

Condensate Receiver Level Cascaded to Steam Flow Control

The necessary equation here is:

$$w_s(s) = \frac{1}{K_{mfs}} K_{chT} G_{chT}(s) H_T(s) \quad (14.33)$$

where

$$\begin{aligned} K_{mfs} &= \text{steam-flow transmitter gain, } \frac{\text{psi}}{\text{lbm/min}} \\ &= 12/(w_s)_{\max} \text{ where } (w_s)_{\max} = \text{steam} \\ &\quad \text{flow meter span, lbm/min} \end{aligned}$$

Note that we assume the flow control loop to be very fast compared with other dynamics. Also, since we have a cascade system, the steam flow transmitter should have a linear relationship between flow and transmitter output. If an orifice flow meter is used, the ΔP transmitter should be followed by a square root extractor.

Base Level Adjusts Feed Flow

$$w_F(s) = \frac{1}{K_{mff}} K_{chB} G_{chB}(s) K_{mbB} H_B(s) \quad (14.34)$$

where

$$\begin{aligned} K_{mff} &= \text{feed flow-meter gain, } \frac{\text{psi}}{\text{lbm/min}} \\ &= 12/(w_F)_{\max} \\ (w_F)_{\max} &= \text{feed flow-meter span, lbm/min} \end{aligned}$$

Reflux Flow Ratioed to Distillate Flow

Let:

$$w_R(s) = K_{RD} G_{RD}(s) w_D(s) \quad (14.35)$$

Physical techniques for accomplishing this are discussed in reference 2.

Bottom-Product Flow Ratioed to Distillate Flow

Let:

$$w_B(s) = K_{BD} G_{BD}(s) w_D(s) \quad (14.36)$$

Closed-Loop Signal-Flow Diagram

The closed-loop signal-flow diagram of Figure 14.6 may now be prepared. To show the relationship between w_D and w_F more clearly, it is redrawn into the form of Figure 14.7. This is a much more complex diagram than that of

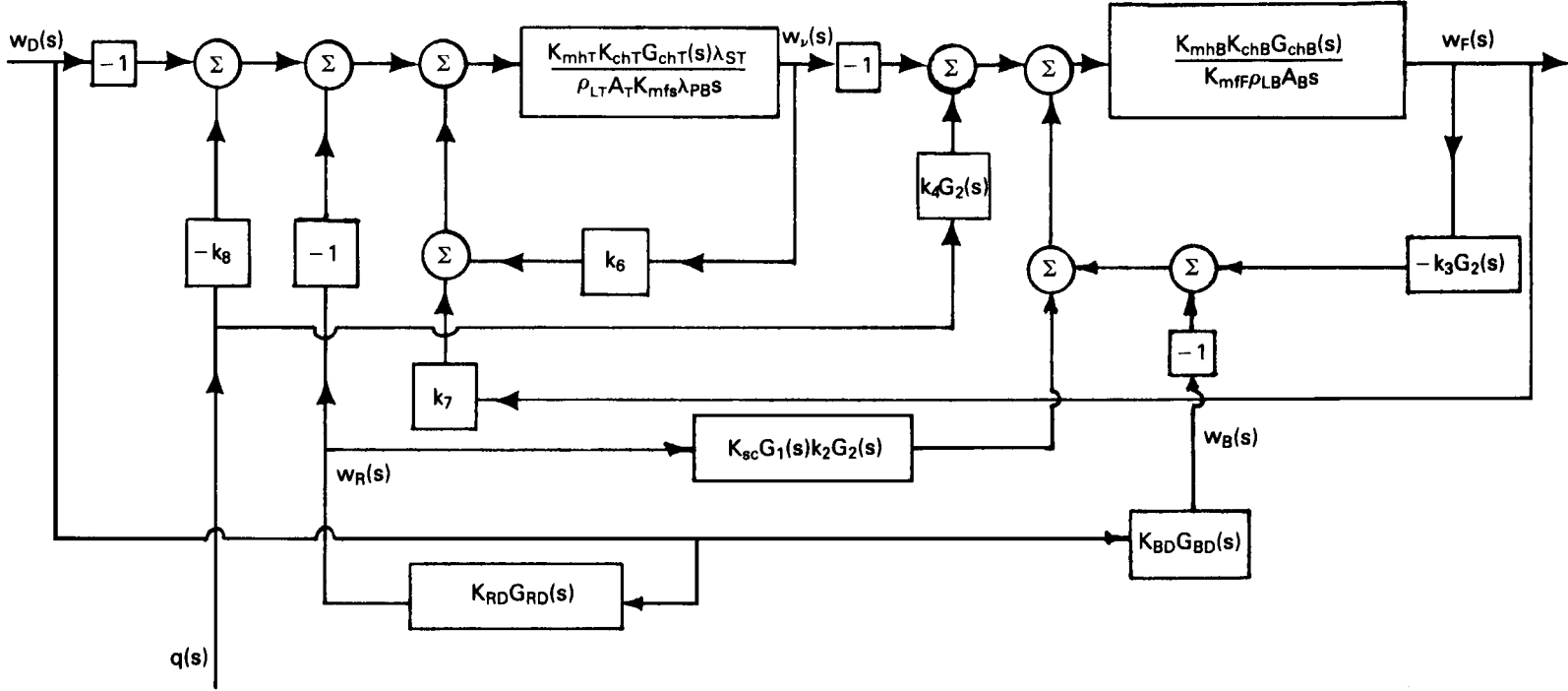


FIGURE 14.7
Rearranged version of Figure 14.6

Figure 14.5. The functions $G_{RD}(s)$ and $G_{BD}(s)$ will have to be chosen with care because of potential difficulties with stability. Also, these two functions must be chosen with primary regard for material-balance control, *not* composition control. This point of view is at variance with that sometimes expressed elsewhere in the literature. Finally, it is probably apparent that conventional “tuning” procedures are essentially useless for a system of this complexity; control functions must be correctly preselected, and control-loop parameters calculated ahead of plant operation.

14.4 MATERIAL-BALANCE CONTROL IN SIDESTREAM DRAWOFF COLUMNS

Let us consider two cases: (1) vapor sidestream, and (2) liquid sidestream. Feed is assumed to enter at its boiling point.

Vapor Sidestream

As an example let us consider a column such as that illustrated in Figure 7.1. This column has a small top-product purge, a small bottom-product purge, and a side product that is most of the feed. Base level adjusts side draw and reflux drum level sets reflux flow. Three flows are ratioed to feed: top product, bottom product, and steam.

About the only new relationship we need is that which defines vapor flow up the column above the point of side draw:

$$w_s(s) = \frac{K_{mhB}}{K_{mfs}} K_{chB} G_{chB}(s) H_B(s) \quad (14.37)$$

= sidestream flow, lbm/min

where

$$K_{mfs} = \text{sidestream flow-meter gain, } \frac{\text{psi}}{\text{lbm/min}}$$

$$= \frac{12}{(w_s)_{\max}}$$

and

$$w_{t-1}(s) = w_v(s) - w_s(s) \quad (14.38)$$

The steam flow is set by ratio to the feed flow:

$$w_s(s) = K_{mfT} K_{R1} G_{R1}(s) \frac{1}{K_{mfs}} w_F(s) \quad (14.39)$$

and

$$w_v(s) = \frac{\lambda_x}{\lambda_p} w_s(s) \quad (14.40)$$

The two remaining ratio controls are defined as follows:

$$w_B(s) = K_{mfF} K_{R3} G_{R3}(s) \frac{1}{K_{mfB}} w_F(s) \quad (14.41)$$

and

$$w_D(s) = K_{mfF} K_{R2} G_{R2}(s) \frac{1}{K_{mfD}} w_F(s) \quad (14.42)$$

The pertinent equations may now be assembled into the form of Figure 14.8.

Liquid Sidestream

In this case we will assume that the liquid side draw is taken from a point above the feed tray. Then we define:

$$\frac{w_{d1}(s)}{w_i(s)} = G_{1a}(s) \quad (14.43)$$

$$w_{d2}(s) = w_{d1}(s) = w_s(s) \quad (14.44)$$

$$\frac{w_{f+1}(s)}{w_{d2}(s)} = G_{1b}(s) \quad (14.45)$$

With the same control scheme in mind, we can prepare the signal flow diagram of Figure 14.9.

14.5 TOP AND BOTTOM LEVEL CONTROL COMBINATIONS

Considerable controversy has existed on the question of whether to have the condensate receiver level adjust the reflux flow or the top-product flow. One well-known author argues strongly for the former. Controversy also exists as to whether it is better to have column-base level control bottom-product flow or the reboiler heating medium, usually steam. Another expert recommends the second. It is probably apparent that we cannot follow both recommendations; at least one of the two levels must control a drawoff flow.

There are many columns operating today with condensate receiver level controlling reflux and base level controlling bottom-product flow. There are other columns in which condensate receiver level adjusts top-product flow while base level manipulates steam flow. How do we choose between them, assuming that we cannot, for some reason, have both levels adjust drawoff flows?

It seems to us that it is largely a matter of convenience. In a superfractionator, for example, the reflux flow may be ten or more times greater than the top-product flow. Inventory in the receiver may be regulated a little more readily by manipulation of the large flow than of the small one. This does not mean that level control via the small flow is either impossible or impractical. It does

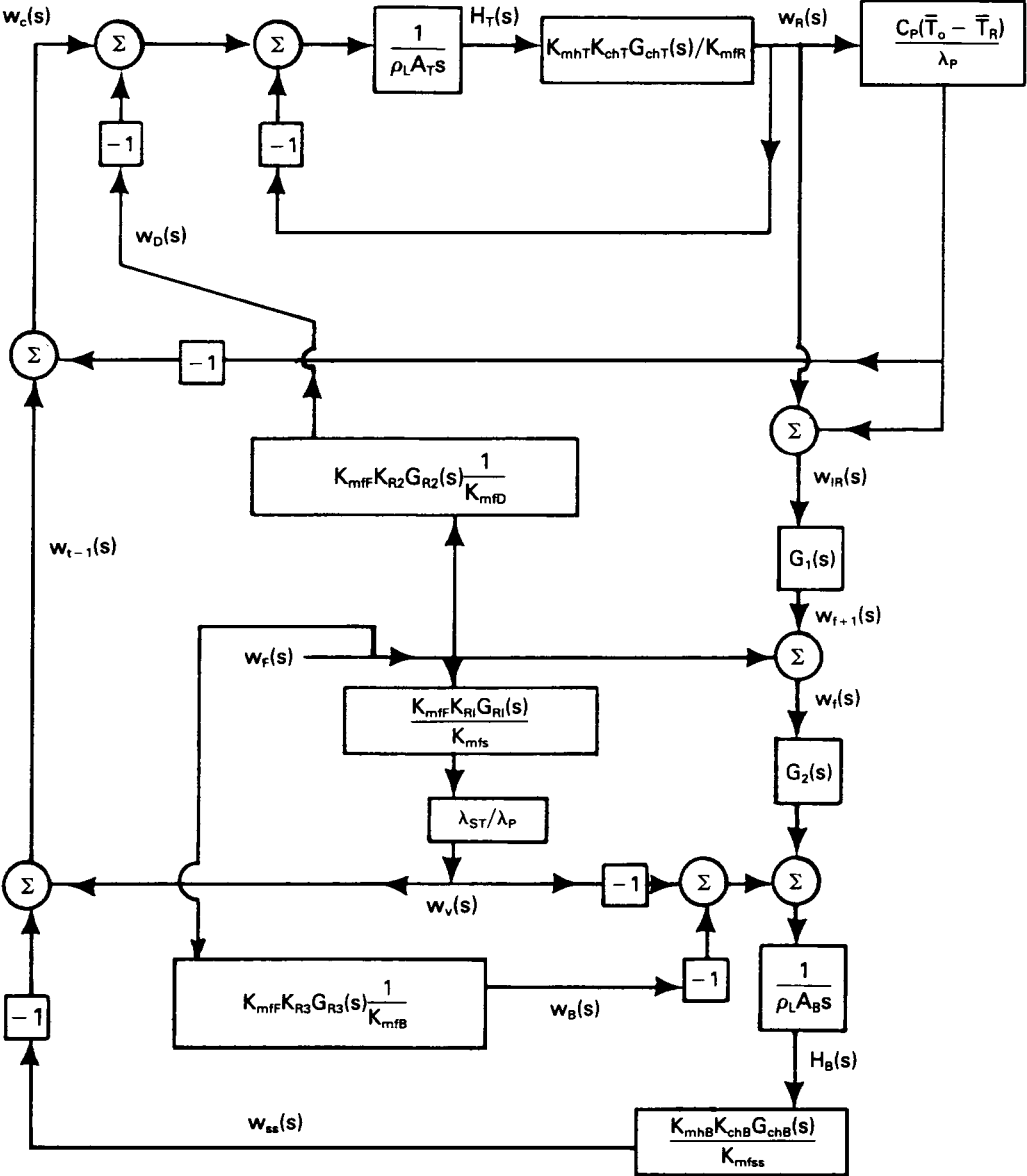


FIGURE 14.8
Material balance signal flow diagram—vapor sidestream drawoff

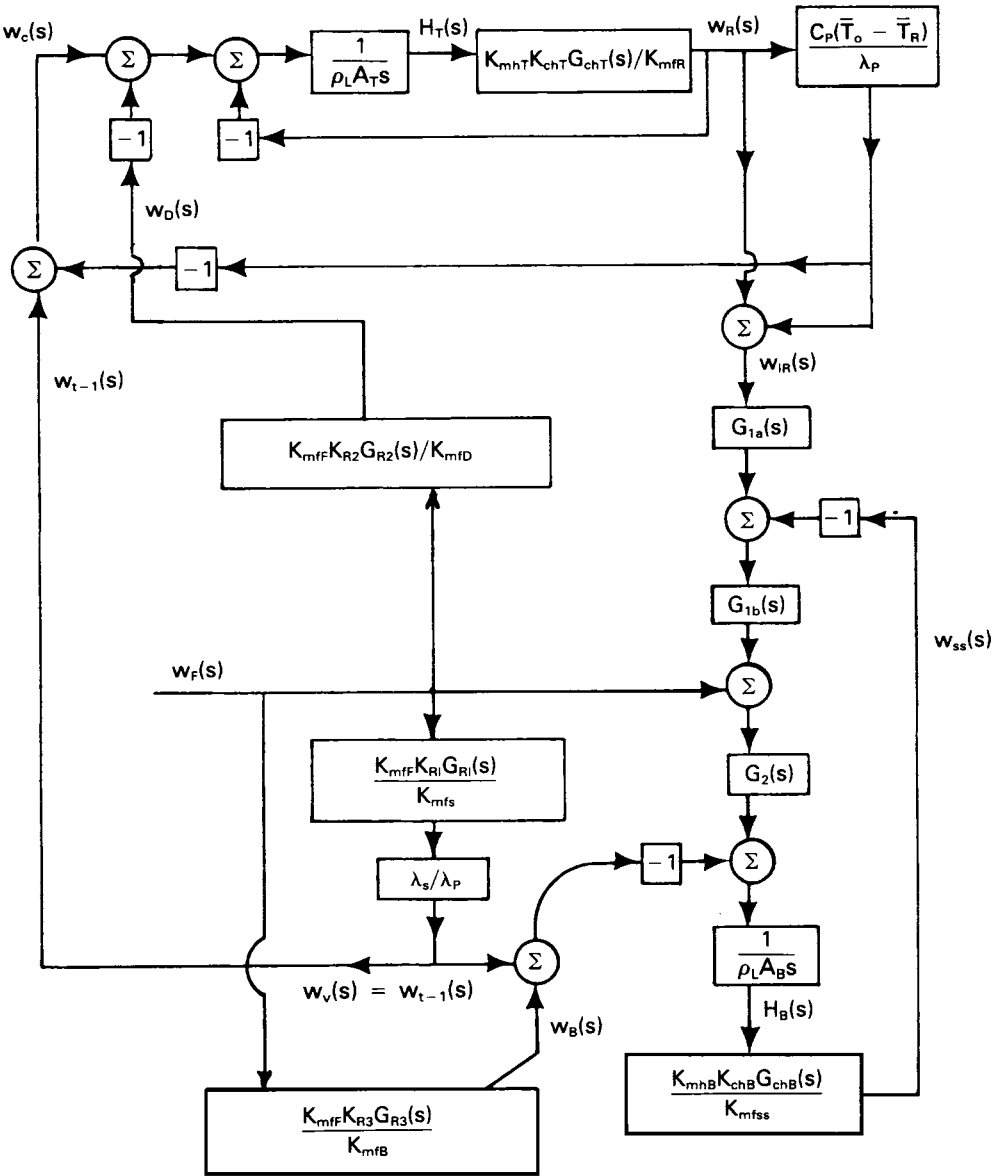


FIGURE 14.9
Material balance signal flow diagram—liquid sidestream drawoff

mean that high- and low-level protection by means of overrides on reflux flow would be needed. This is discussed in Chapters 9 and 16.

It is sometimes argued that where reflux flow is much greater than top-product flow, one may control top composition more easily by adjusting top-product flow than by adjusting reflux flow. Actually a little algebra will show that there is not much difference, and that the difference is against the argument rather than in favor of it. If, for example, a change in feed rate or feed composition changes overhead composition, there will be a certain change in reflux flow and another change in top-product flow required to restore the top composition. These two required changes are the same in the steady state regardless of which variable is manipulated to control top composition. Composition control via distillate (top product) has the disadvantage that no change in composition takes place until the reflux flow changes. Since reflux is controlled by level, the dynamics of the level control loop appear in the composition control loop. This means, generally, that we cannot use averaging level control; we must design for tight level control. For this reason we normally prefer to control composition via reflux.

A similar line of reasoning may be followed at the base of the column, and leads to the conclusion that we would normally prefer to control base composition by manipulating boilup. Controlling base level by steam has another disadvantage if a thermosyphon reboiler is used; interchange of inventory between column base and reboiler sometimes leads to severe dynamic problems. This is discussed in Chapters 4, 15, and 16.

15

Condenser and Reboiler Dynamics

15.1 LIQUID-COOLED CONDENSERS WITH NO CONDENSATE HOLDUP

as mentioned in Chapter 3, we find in the chemical and petroleum industries two principal types of liquid-cooled condensers: (1) the horizontal type with vapor on the shell side and coolant in the tubes, and (2) the vertical design with vapor in the tubes and coolant on the shell side. We can also think of condensers in terms of whether the coolant goes through just once (no axial mixing) or is recirculated to achieve good axial mixing. A condenser with the latter type of cooling is said to have “tempered” cooling.

Condenser with No Axial Mixing of Coolant

Once-through coolant is by far the most common choice. An approximate analysis for a condenser that has a single pass on the coolant side is presented in Chapter 24 of reference 1, and will not be repeated here. It involves the following simplifying assumptions:

1. Heat storage in the heat exchanger metal is negligible.
2. Subcooling is negligible.
3. Mean temperature difference is arithmetic.

Although these reduce the complexity somewhat, we are still left with the job of solving partial differential equations. Perhaps the easiest to read descriptions of their solutions are those by Hempel² and by Gould.³ A much simpler model, largely empirical, has been proposed by Thal-Larsen.⁴ Since most liquid-cooled condensers are fairly fast with time constants in the range of 10–60 seconds, we will not pursue their dynamic equations further. For important applications, where subcooling may be of concern, one should probably resort to simulation.

The literature on subcooled condensers is very sparse; one paper has been published by Luyben, Archambault, and Jauffret,⁵ and another by Tyreus.⁶

If the sensible heat load of subcooling is not too large compared with that of the condensing heat load (and this is usually the case), the following static gains may be derived:

$$\frac{\partial w_c}{\partial w_A} = \frac{(\bar{T}_c - \bar{T}_{A1}) \left[2 \bar{U}_c A_c^2 c_A + 4 \bar{w}_A^2 c_A^2 A_c \frac{\partial U_c}{\partial w_A} \right]}{\lambda_p (\bar{U}_c A_c + 2 \bar{w}_A c_A)^2} \quad (15.1)$$

$$\frac{\partial w_c}{\partial T_{A1}} = \frac{-2 \bar{U}_c A_c \bar{w}_A c_A}{\lambda_p (\bar{U}_c A_c + 2 \bar{w}_A c_A)} \quad (15.2)$$

$$\frac{\partial w_c}{\partial T_c} = \frac{2 \bar{U}_c A_c \bar{w}_A c_A}{\lambda_p (\bar{U}_c A_c + 2 \bar{w}_A c_A)} \quad (15.3)$$

$$\frac{\partial T_{AO}}{\partial T_c} = \frac{2 \bar{U}_c A_c}{\bar{U}_c A_c + 2 \bar{w}_A c_A} \quad (15.4)$$

$$\frac{\partial T_{AO}}{\partial T_{A1}} = \frac{-(\bar{U}_c A_c - 2 \bar{w}_A c_A)}{\bar{U}_c A_c + 2 \bar{w}_A c_A} \quad (15.5)$$

$$\frac{\partial T_{AO}}{\partial w_A} = \frac{4 \bar{w}_A c_A A_c \frac{\partial U_c}{\partial w_A} (\bar{T}_c - \bar{T}_{A1}) - 4 \bar{U}_c A_c c_A (\bar{T}_c - \bar{T}_{A1})}{(\bar{U}_c A_c + 2 \bar{w}_A c_A)^2} \quad (15.6)$$

where

T_c = process condensing temperature, °K

T_{A1} = coolant inlet temperature, °K

T_{AO} = coolant exit temperature, °K

c_A = coolant specific heat, pcu/lbm °C

U_c = condenser heat-transfer coefficient, $\frac{\text{pcu/sec}}{^\circ\text{C ft}^2}$

w_c = rate of condensation, lbm/sec

w_A = coolant flow rate, lbm/sec

λ_p = latent heat, pcu/lbm, of process vapor

A_c = condenser heat-transfer area, ft²

Condenser with Well-Mixed Coolant

Qualitatively the condenser with well-mixed (tempered) coolant is discussed in Chapter 3, Section 9 (see Figure 3.19). A mathematical analysis of its dynamics is given in Chapter 24 of reference 1. We will not repeat it here, but it leads to the following results (simplifying assumptions are the same as in the previous section):

$$\left[\frac{w_c(s)}{w_A(s)} \right]_{OL} = \frac{c_A (\bar{T}_A - \bar{T}_{A1}) K_2}{\tau_2 s + 1} \quad (15.7)$$

$$\left[\frac{w_c(s)}{T_{A1}(s)} \right]_{OL} = \frac{-\bar{w}_A c_A K_2}{\tau_2 s + 1} \quad (15.8)$$

$$\left[\frac{w_c(s)}{T_c(s)} \right]_{OL} = \frac{\bar{w}_A c_A K_2 (\tau_1 s + 1)}{\tau_2 s + 1} \quad (15.9)$$

where

$$\tau_1 = \frac{W_A}{\bar{w}_A}$$

T_A = average bulk temperature, °K, of coolant

W_A = coolant holdup, lbm

$$\tau_2 = \frac{W_A / \bar{w}_A}{\frac{\bar{U}_c A_c}{\bar{w}_A c_A} + 1}$$

$$K_2 = \frac{\bar{U}_c A_c / \lambda_p}{\bar{U}_c A_c + \bar{w}_A c_A}$$

The subscript *OL* means open loop. Also:

$$\left[\frac{T_A(s)}{T_c(s)} \right]_{OL} = \frac{\bar{U}_c A_c}{\bar{U}_c A_c + \bar{w}_A c_A} \times \frac{1}{\tau_2 s + 1} \quad (15.10)$$

$$\left[\frac{T_A(s)}{w_A(s)} \right]_{OL} = \frac{-c_A (\bar{T}_A - \bar{T}_{A1})}{\bar{U}_c A_c + \bar{w}_A c_A} \times \frac{1}{\tau_2 s + 1} \quad (15.11)$$

$$\left[\frac{T_A(s)}{T_{A1}(s)} \right]_{OL} = \frac{\bar{w}_A c_A}{\bar{U}_c A_c + \bar{w}_A c_A} \times \frac{1}{\tau_2 s + 1} \quad (15.12)$$

The first-order dynamics of this type of condenser make it much easier to control than the condenser with once-through coolant.

15.2 FLOODED CONDENSERS—OPEN-LOOP DYNAMICS

Most flooded condensers are of the horizontal type with vapor on the shell side and coolant in the tubes as shown in Figure 15.1 (see also Figure 3.1). As the liquid level in the shell varies, so do heat-transfer area and rate of condensation. We would like to find out how condensing pressure and rate of condensation are affected by rate of vapor flow into the condenser, back pressure downstream of the vent valve, and rate of removal of liquid from the condenser shell. We will make several simplifying assumptions:

1. The sensible heat load is small enough in comparison with the latent heat load that it may be neglected.
2. Submerged heat-transfer area, A_s , is proportional to liquid level above the bottom of the lowest tube. The change in condensing area, A_c , is then the negative of the change in A_s . This assumption is not bad if there are many tubes and if they are not "layered." To make this assumption valid may require that in some cases the tube bundle be slightly rotated about its axis.
3. For the time being, we will assume that the vent valve position is fixed.
4. Heat storage in the heat exchanger metal may be neglected.

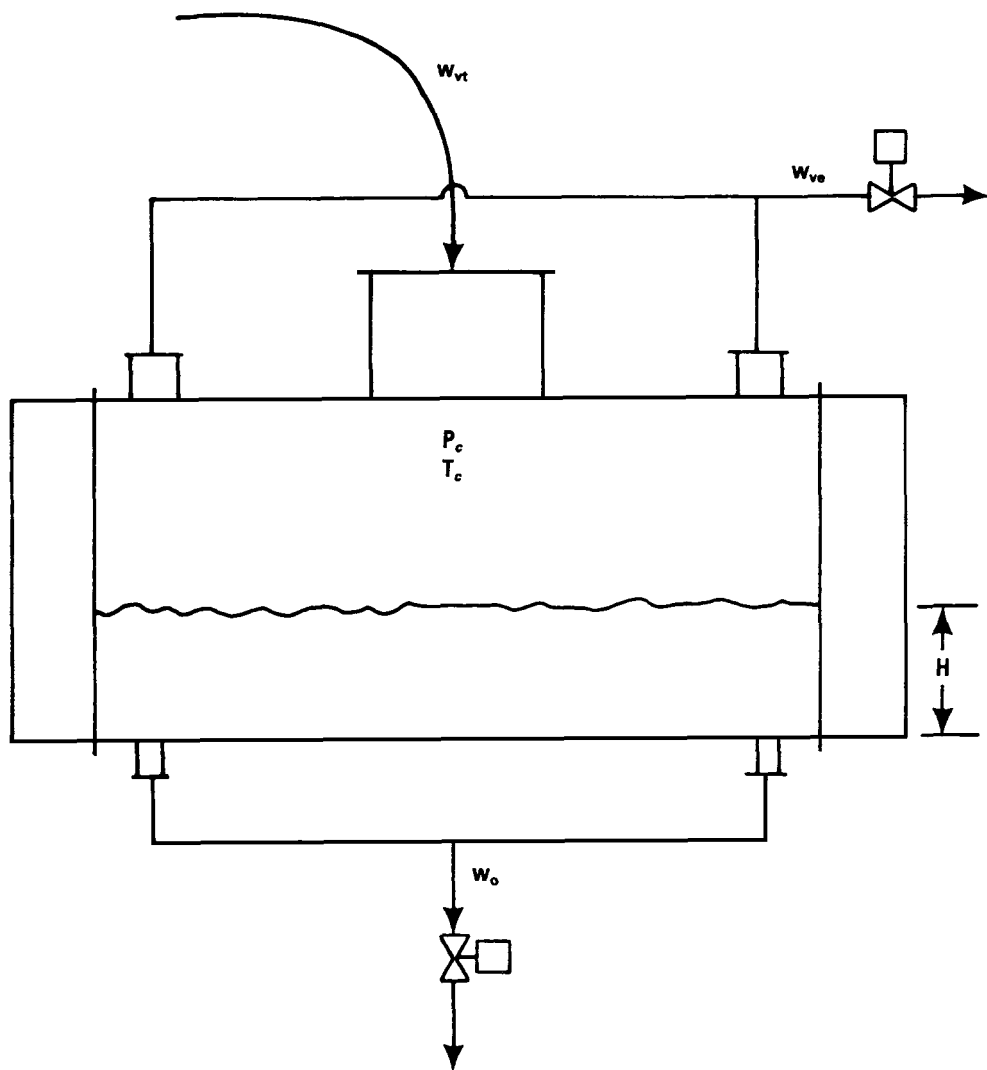


FIGURE 15.1

Horizontal condenser with coolant in tubes and partially flooded on shell side

We may now write the following equations, some in the time domain and some in the s domain:

$$q_c(t) = U_c A_c(t) \Delta T(t) \quad (15.13)$$

$$w_c(t) = \frac{q_c(t)}{\lambda_p} \quad (15.14)$$

$$\frac{w_c(s) - w_o(s)}{s} = W_{SH}(s) \quad (15.15)$$

$$w_{vt}(t) - w_c(t) = w_{ve}(t) \quad (15.16)$$

$$A_s(s) \cong \frac{\partial A_s}{\partial W_{SH}} W_{SH}(s) \quad (15.17)$$

$$A_c(s) = -A_s(s) \quad (15.18)$$

$$\Delta T(s) = T_c(s) - T_A(s) \quad (15.19)$$

$$T_c(s) = \frac{\partial T_c}{\partial P_c} P_c(s) \quad (15.20)$$

$$\frac{P_c(s) - P_R(s)}{R_p} = \frac{w_{ve}(s)}{\rho_p} \quad (15.21)$$

$$P_c(s) = \sum Q(s)/(V/\bar{P}_c)s \quad (15.21a)$$

where

- V = system vapor volume, ft^3
- q_c = rate of heat transfer, pcu/sec
- U_c = condenser heat-transfer coefficient, $\frac{\text{pcu/sec}}{\text{ft}^2 \text{ } ^\circ\text{C}}$
- A_c = condensing heat-transfer area, ft^2
- T = $^\circ\text{C}$ or $^\circ\text{K}$
- w_c = lbm/sec process vapor condensed
- λ_p = latent heat of vaporization, pcu/lbm , of condensing process vapor
- w_o = liquid outflow, lbm/sec
- w_{vt} = vapor inflow, lbm/sec
- w_{ve} = vapor outflow, lbm/sec
- A_s = submerged heat-transfer area, ft^2
- T_c = condensing temperature, $^\circ\text{C}$ or $^\circ\text{K}$
- T_A = average coolant temperature, $^\circ\text{C}$ or $^\circ\text{K}$
- P_c = vapor space pressure, lbf/ft^2
- P_R = pressure, lbf/ft^2 , downstream of vent valve
- R_p = vent valve resistance, lbf sec/ft^5
- W_{SH} = pounds of liquid in shell between bottom of lowest tube and top of highest tube
- A_{TC} = total heat-transfer area, ft^2 , of condenser
- Q = flow, ft^3/sec

Upon Laplace transforming equations (15.13), (15.14), and (15.16), we obtain:

$$q_c(s) = U_c \bar{A}_c \Delta T(s) + U_c \bar{\Delta T} A_C(s) \quad (15.22)$$

$$w_c(s) = \frac{q_c(s)}{\lambda_p} \quad (15.23)$$

$$w_r(s) - w_c(s) = w_{re}(s) \quad (15.24)$$

Response, $P_c(s)$, to Various Inputs

To find the response, $P_c(s)$, to various inputs, we combine equations (15.13) through (15.24) into the signal flow diagram of Figure 15.2. The first reduction of this is given in Figure 15.3 and the final reduction in Figure 15.4. From this we may write by inspection:

$$P_c(s) = \left[\frac{\frac{R_v/\rho_v}{\alpha}(s + \alpha)}{\frac{\bar{V}}{\bar{P}_c} R_v s^2 + \frac{1 + \alpha R_v \frac{\bar{V}}{\bar{P}_c} + K_{CO}}{\alpha} s + 1} \right] \times \left[\frac{-\alpha}{s + \alpha} w_o(s) + \frac{\bar{U}_c A_c s / \lambda_p}{s + \alpha} T_A(s) + \frac{\rho_v}{R_v} P_R(s) + w_{re}(s) \right] \quad (15.25)$$

where

$$\alpha = \frac{\partial A_s}{\partial W_{SH}} \frac{U_c \bar{\Delta T}}{\lambda_p}$$

$$K_{CO} = \frac{R_v/\rho_v}{\lambda_p} \frac{\partial T_C}{\partial P_c} U_c A_c$$

The first term on the right-hand side of equation (15.25) may be written:

$$\frac{K_{FC}(s + \alpha)}{\tau_Q^2 s^2 + 2\zeta \tau_Q s + 1}$$

For all such condensers we have studied, the denominator has a large damping ratio so that the quadratic may be factored into two terms, one of which has a typical time constant of 1.5–3 minutes, while the other is only a few seconds. The above then reduces to:

$$\frac{K_{FC}(s + \alpha)}{\tau_{FC} s + 1}$$

where

$$\tau_{FC} = \frac{1 + \alpha R_v \frac{\bar{V}}{\bar{P}_c} + K_{CO}}{\alpha}$$

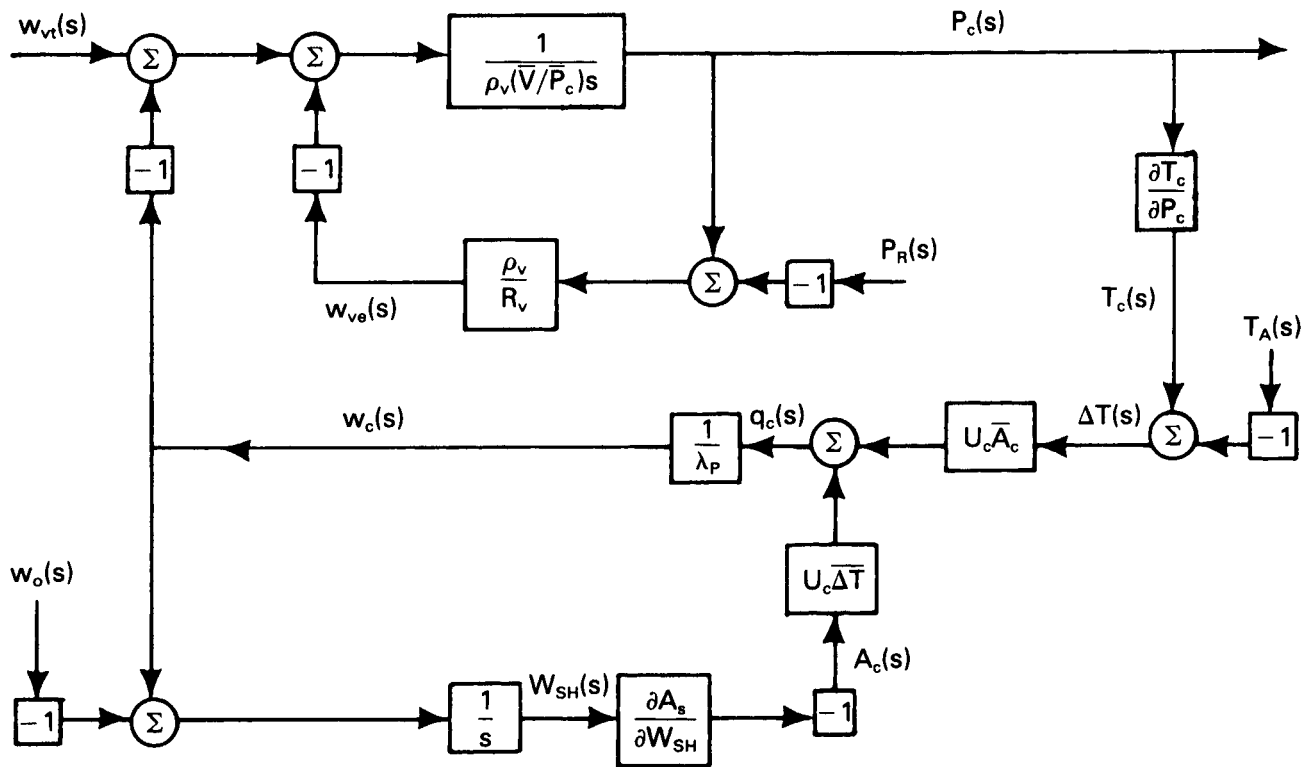


FIGURE 15.2
First signal flow diagram for P_c of flooded condenser

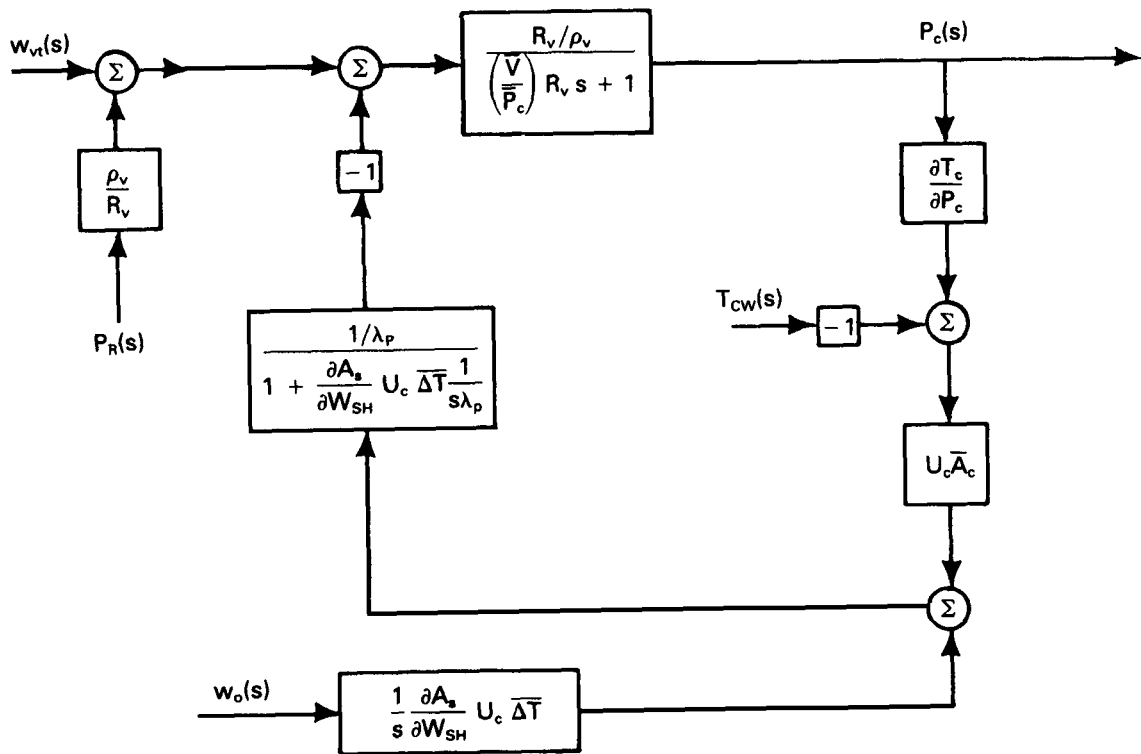


FIGURE 15.3
First reduction of signal flow diagram of figure 15.2

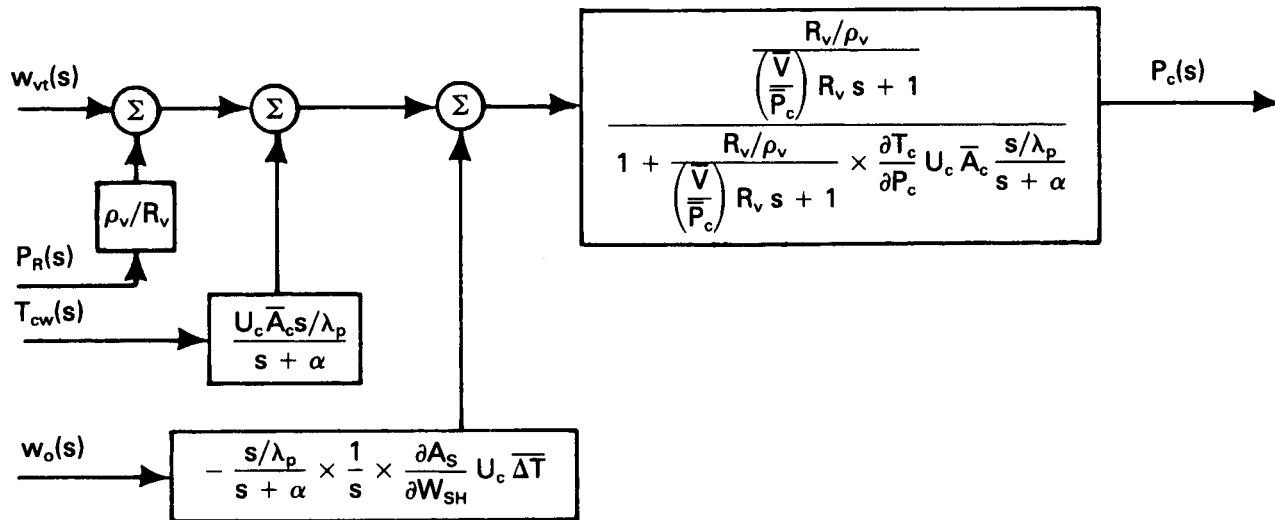


FIGURE 15.4
Final signal flow diagram for P_c of flooded condenser

Case Where R_v Approaches Infinity

For the case where R_v becomes very large (very little vent gas flow), the last equation reduces to:

$$P_c(s) = \frac{(1/\rho_v)(s + \alpha)}{\left(\alpha \bar{V}/\bar{P}_c + \frac{U_c \bar{A}_c}{\lambda_p \rho_v} \frac{\partial T_c}{\partial P_c} \right) s \left(\frac{\bar{V}/\bar{P}_c}{\alpha \bar{V}/\bar{P}_c + \frac{U_c \bar{A}_c}{\lambda_p \rho_v} \frac{\partial T_c}{\partial P_c}} s + 1 \right)} \times \left[\frac{-\alpha}{s + \alpha} w_o(s) + \frac{U \bar{A}_c / \lambda_p}{s + \alpha} T_A(s) + w_w(s) \right] \quad (15.26)$$

The first term on the right may be written:

$$\frac{K'_{FC}(s + \alpha)}{s(\tau'_{FC}s + 1)}$$

where

$$K'_{FC} = \frac{1/\rho_v}{\alpha \bar{V}/\bar{P}_c + \frac{U_c \bar{A}_c}{\lambda_p \rho_v} \frac{\partial T_c}{\partial P_c}}$$

and

$$\tau'_{FC} = \frac{\bar{V}/\bar{P}_c}{\alpha \bar{V}/\bar{P}_c + \frac{U_c \bar{A}_c}{\lambda_p \rho_v} \frac{\partial T_c}{\partial P_c}}$$

Response, $w_c(s)$, to Various Inputs

The signal flow diagram of Figure 15.2 can be redrawn to show $w_c(s)$ as an output as shown in Figure 15.5. This can then be reduced to the form of Figure 15.6, from which we can write by inspection:

$$w_c(s) = \Omega(s) \left(\frac{\alpha \lambda_p}{s} w_o(s) - U_c \bar{A}_c T_A(s) + \left[\frac{(R_v/\rho_v) \bar{U}_c A_c \frac{\partial T_c}{\partial P_c}}{\frac{\bar{V}}{\bar{P}_c} R_v s + 1} \right] \times \left[w_w(s) + \frac{\rho_v}{R_v} P_R(s) \right] \right) \quad (15.27)$$

where

$$\Omega(s) = \frac{\frac{s}{\alpha \lambda_p} \left(\frac{\bar{V}}{\bar{P}_c} R_v s + 1 \right)}{\left[\frac{\bar{V}}{\bar{P}_c} R_v \frac{s^2}{\alpha} + \frac{1 + \alpha \frac{\bar{V}}{\bar{P}_c} R_v + K_{CO}}{\alpha} s + 1 \right]} = \frac{K_{FC}'' \left(\frac{\bar{V}}{\bar{P}_c} R_v s + 1 \right)}{\tau_Q^2 s^2 + 2\zeta \tau_Q s + 1} \quad (15.27a)$$

Case Where R_v Approaches Infinity

Again, as R_v becomes very large, the preceding equation becomes simpler:

$$w_c(s) = \left[\frac{(\bar{V}/\bar{P}_c \lambda_p)}{\alpha \bar{V}/\bar{P}_c + \frac{U_c \bar{A}_c}{\rho_v \lambda_p} \frac{\partial T_c}{\partial P_c}} \times \frac{s}{\frac{\bar{V}/\bar{P}_c}{\alpha \bar{V}/\bar{P}_c + \frac{U_c \bar{A}_c}{\rho_v \lambda_p} \frac{\partial T_c}{\partial P_c}} s + 1} \right] \times \left[\frac{\alpha \lambda_p}{s} w_o(s) - U_c \bar{A}_c T_A(s) + \frac{1}{\frac{\rho_v \bar{V}}{\bar{P}_c} s} w_m(s) \right] \quad (15.28)$$

15.3 REBOILERS—OPEN-LOOP DYNAMICS

We wish to make an analysis of column bases with associated reboilers where there is significant liquid holdup. The analysis should take into account the temperature of the entering liquid and the sensible heat effect of the liquid mass. Such an analysis also applies to vaporizers with associated separators or knockout drums.

Heat-Transfer Dynamics

The combination of thermosyphon reboiler (or any high circulation rate reboiler) and column base or separator may be represented as shown schematically in Figure 15.7. The various equations may then be written as follows.

Heat Balance

$$w_i(t) c_p T_i(t) + q_T(t) - (w_{BU}(t) [\lambda_p + c_p T_{BU}(t)] + w_B(t) c_p T_{BU}(t)) = \frac{dW_B(t) c_p T_{BU}(t)}{dt} \quad (15.29)$$

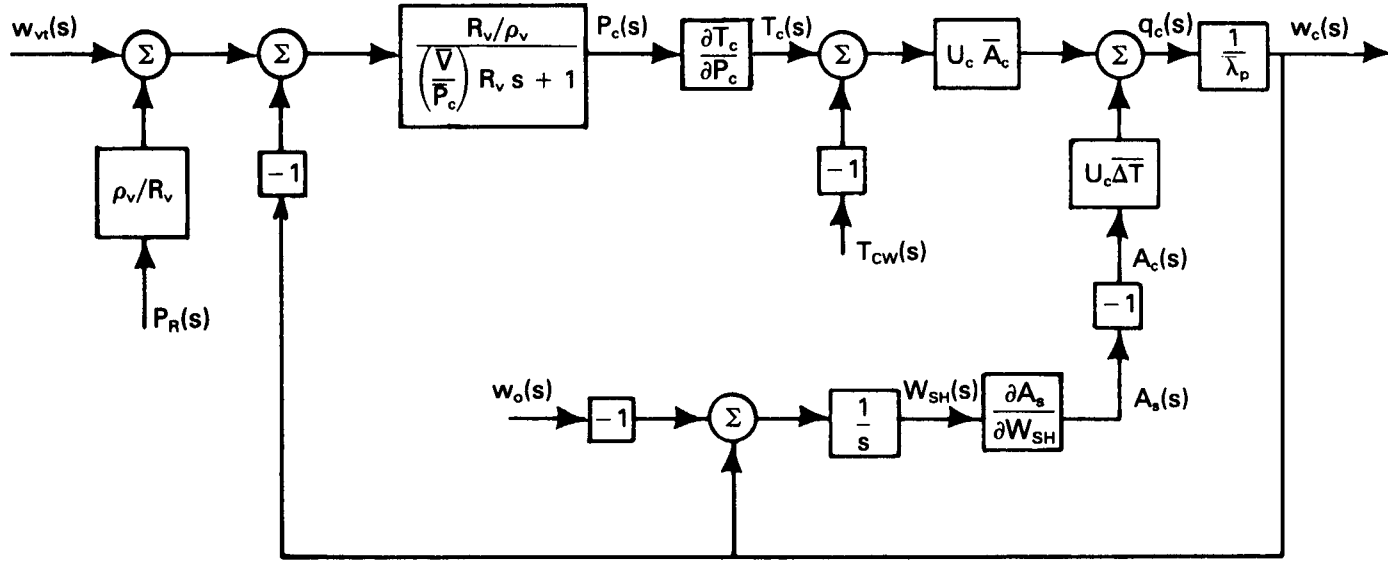


FIGURE 15.5
First signal flow diagram for w_c of flooded condenser

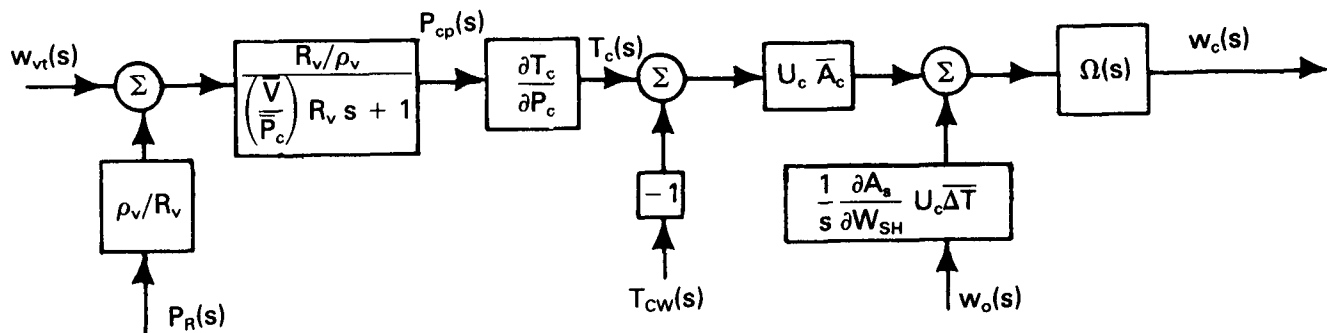


FIGURE 15.6
Reduced signal flow diagram for w_c of flooded condenser

By using perturbation techniques and Laplace transforming this equation we get:

$$c_p \bar{T}_i w_i(s) + c_p \bar{w}_i T_i(s) + q_T(s) - \lambda_p w_{BU}(s) - c_p \bar{w}_{BU} T_{BU}(s) - c_p \bar{T}_{BU} w_{BU}(s) \\ - c_p \bar{w}_B T_{BU}(s) - c_p \bar{T}_{BU} w_B(s) = c_p s [\bar{T}_{BU} W_B(s) + \bar{W}_B T_{BU}(s)] \quad (15.30)$$

Material Balance

$$\int [w_i(t) - w_{BU}(t) - w_B(t)] dt = W_B(t) \quad (15.31)$$

or

$$[w_i(s) - w_{BU}(s) - w_B(s)] \frac{1}{s} = W_B(s) \quad (15.32)$$

Steam-Side Dynamics

The following are taken from Chapter 25 of reference 1:

$$T_{cs}(s) = \frac{\partial T_{cs}}{\partial P_{cs}} P_{cs}(s) \quad (15.33)$$

$$P_{cs}(s) = \frac{Q_i(s) - Q_o(s)}{C_R s} \quad (15.34)$$

$$Q_i(s) = \frac{\partial Q_v}{\partial P_s} P_s(s) + \frac{\partial Q_v}{\partial P_{cs}} P_{cs}(s) + \frac{\partial Q_v}{\partial X_v} X_v(s) \quad (15.35)$$

$$Q_o(s) = \frac{q_T(s)}{\lambda_{re} \rho_{re}} \quad (15.36)$$

$$q_T(s) = U_R A_R [T_{cs}(s) - T_{BU}(s)] \quad (15.37)$$

(Note that heat storage of the reboiler metal is neglected.)

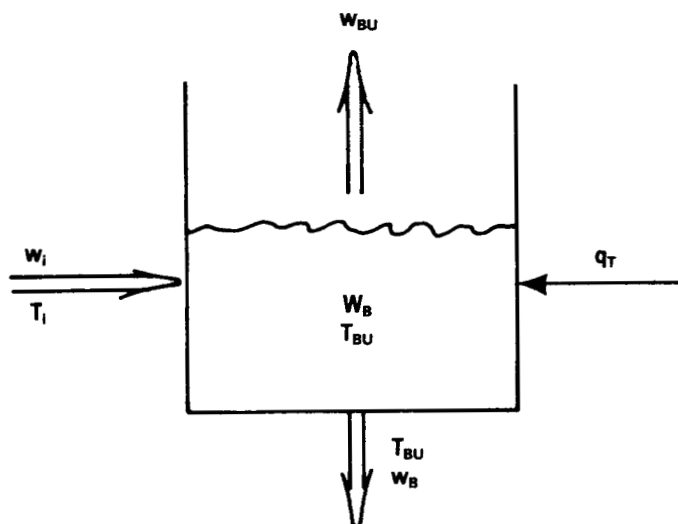


FIGURE 15.7
Schematic representation of column base and reboiler holdup

In these equations:

$T_\alpha(s)$ = steam condensing temperature, °C

$P_\alpha(s)$ = reboiler shell pressure, lbf/ft²

Q_s = steam flow rate, ft³/sec

q_T = heat transfer, pcu/sec

U_R = reboiler heat-transfer coefficient, $\frac{\text{pcu}/^\circ\text{C}}{\text{ft}^2 \text{ sec}}$

A_R = heat-transfer area, ft²

T_{BU} = boiling temperature of process fluid, °C

Q_c = rate of steam condensation, ft³/sec

λ_{st} = steam latent heat of condensation, pcu/lbm

ρ_{st} = steam density lbm/ft³ at P_{CS} and T_{CS}

C_R = acoustic capacitance of reboiler shell, ft⁵/lbf
= V_s/\bar{P}_α where V_s = reboiler shell volume, ft³

P_s = steam supply pressure, lbf/ft², upstream of control valve

X_v = valve stem position

The terms $\partial Q_v/\partial P_s$, $\partial Q_v/\partial P_\alpha$, and $\partial Q_v/\partial X_v$ may be evaluated by the methods of Chapter 15 of reference 1.

The column-base pressure dynamics may be represented by:

$$P_B(s) = \frac{1}{\rho_{BU}} Z_{col}(s) w_{BU}(s) \quad (15.38)$$

where

Z_{col} = column acoustic impedance, looking up from the base, lbf sec/ft⁵

ρ_{BU} = density of vapor boilup, lbm/ft³

w_{BU} = rate of boilup, lbm/min

P_B = column-base pressure, lbf/ft²

From the preceding equations we can prepare the signal flow diagram of Figure 15.8. This can be partially reduced as shown on Figure 15.9 where three new functions are defined:

$$\gamma(s) = \frac{\frac{1}{\lambda_p + c_p \bar{T}_{BU}}}{1 + \left[\frac{c_p (s \bar{W}_B + \bar{w}_{BU} + \bar{w}_{BP})}{\lambda_p + c_p \bar{T}_{BU}} \right] \frac{\partial T_\alpha}{\partial P_\alpha} \frac{Z_{col}}{\rho_{BU}}} \quad (15.39)$$

$$\phi(s) = \frac{U_R A_R}{1 + U_R A_R \gamma(s) \frac{\partial T_\alpha}{\partial P_\alpha} \frac{Z_{col}(s)}{\rho_{BU}}} \quad (15.40)$$

$$\Delta(s) = \frac{U_R A_R \gamma(s) \frac{\partial T_\alpha}{\partial P_\alpha} \frac{Z_{col}(s)}{\rho_{BU}}}{1 + U_R A_R \gamma(s) \frac{\partial T_\alpha}{\partial P_\alpha} \frac{Z_{col}(s)}{\rho_{BU}}} \times \frac{1}{\rho_{st} \lambda_{st}} \quad (15.41)$$

For all reboilers examined to date, $\Delta(s)$ has been so small, both statically and dynamically, as to be negligible. It therefore will be omitted in the remainder of this book. It is advisable, however, to calculate $\Delta(s)$ for any new system as a check.

We have also found that the sensible heat effect of the liquid mass in the column base or separator is small. Heat-transfer lags are typically only several seconds; vapor flow from the separator follows steam flow almost instantaneously.

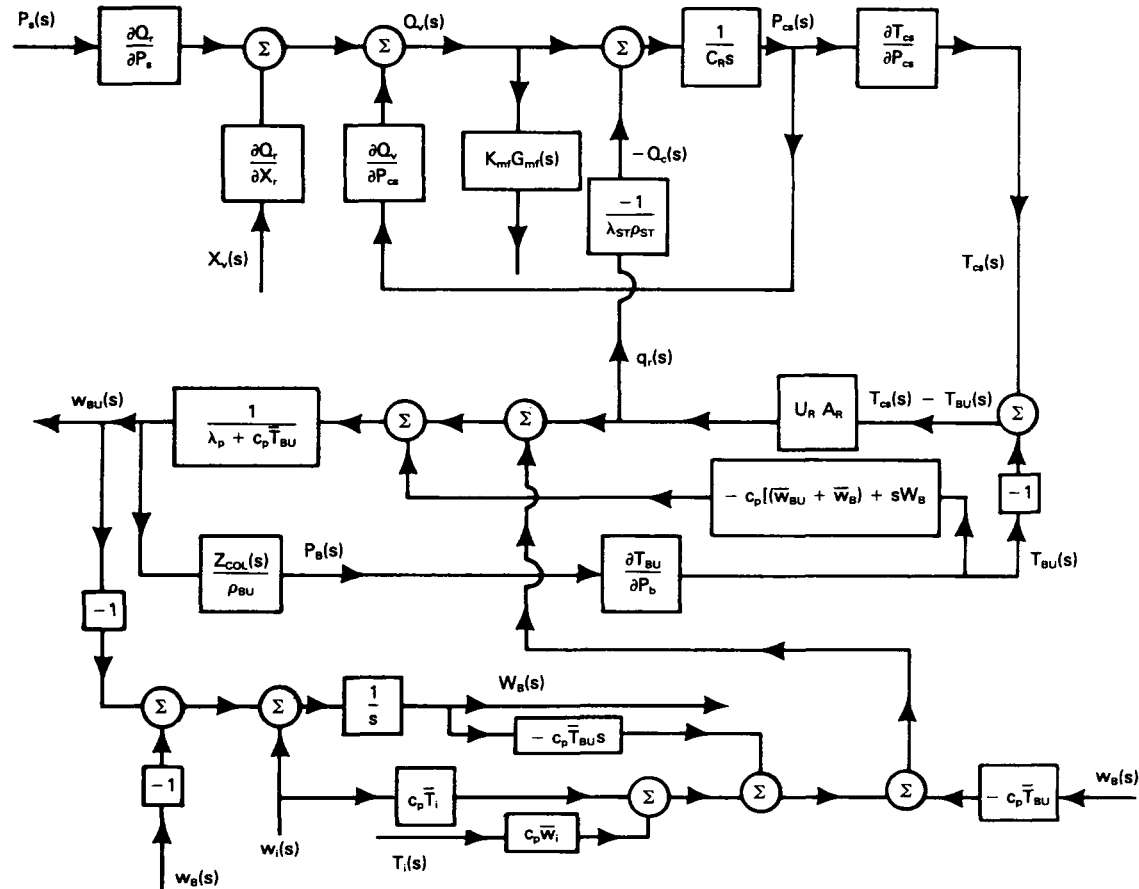


FIGURE 15.8
Preliminary signal flow diagram for heat transfer dynamics

Base Level Control Cascaded to Steam Flow Control

We assume here that averaging level control is desired. We may then prepare the signal flow diagram of Figure 15.10. Note that:

$K_{cf} G_{cf}(s)$ = flow controller transfer function

K_{mh} = liquid-level transmitter gain

$K_{cb} G_{cb}(s)$ = level controller transfer function

A_B = cross-sectional area of column base, ft^2

ρ_L = liquid density, lbm/ft^3

Since T_i varies slowly or not at all, $T_i(s) = 0$.

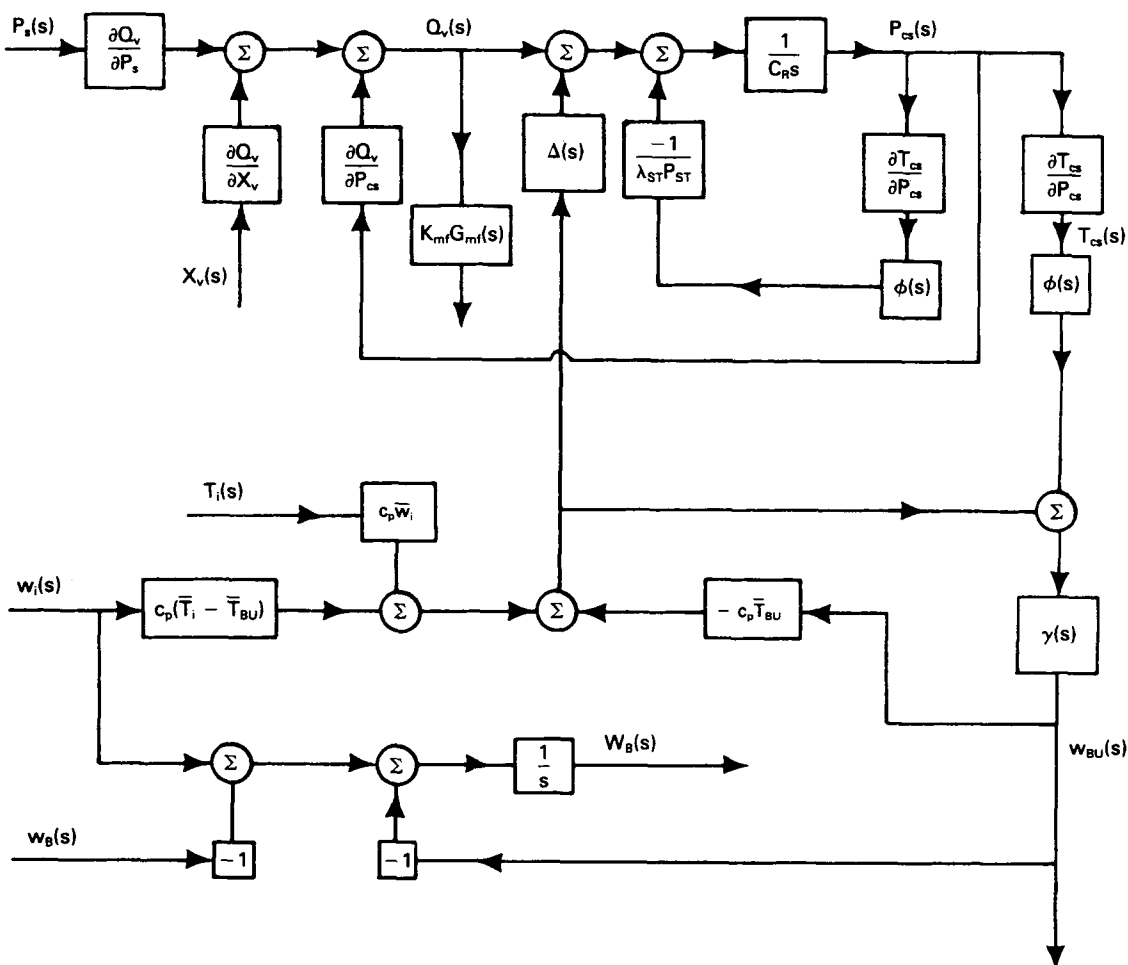


FIGURE 15.9
Partial reduction of figure 15.8

Noncritical Versus Critical Steam Flow

If steam flow is critical, then:

$$\frac{1}{1 - \psi(s) \frac{\partial Q_v}{\partial P_{cs}}} = 1 \tag{15.42}$$

since

$$\frac{\partial Q_v}{\partial P_{cs}} = 0$$

Flow control loop gain and dynamics are determined entirely by the instrument characteristics. Note that:

$$\psi(s) = \frac{\frac{1}{C_R s}}{1 + \frac{1}{C_R s} \frac{1}{\lambda_{\pi} \rho_{\pi}} \phi(s) \frac{\partial T_{cs}}{\partial P_{cs}}} \tag{15.43}$$

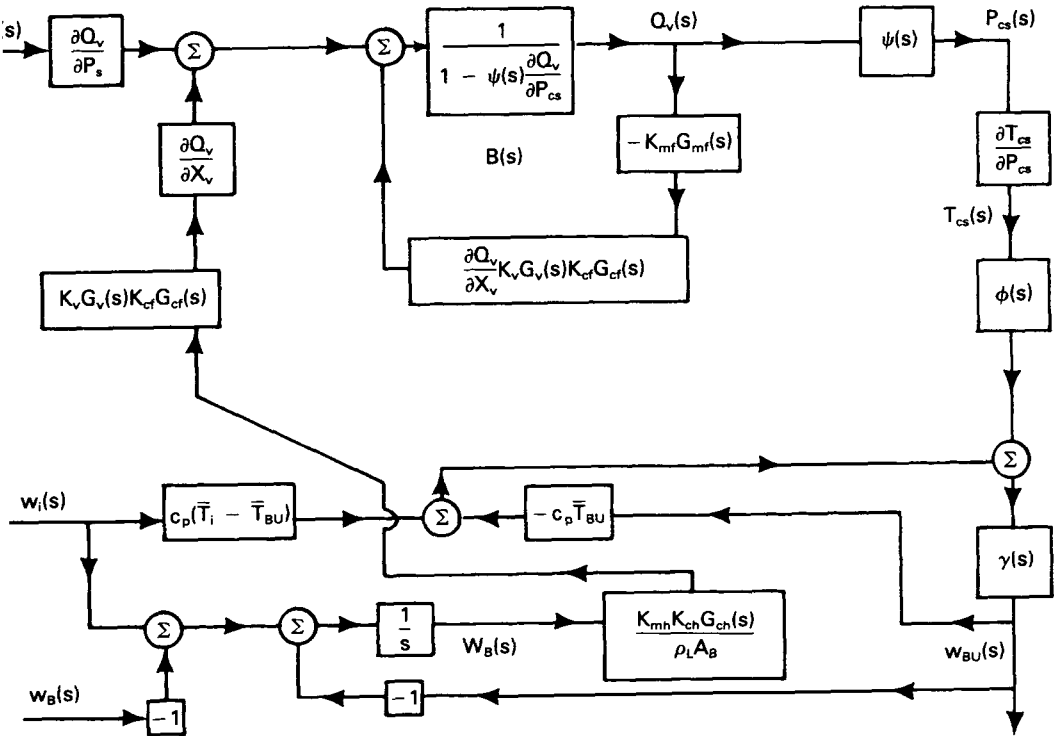


FIGURE 15.10
Signal flow diagram for base level control
cascaded to steam flow control

If steam flow is noncritical,

$$\frac{1}{1 - \psi(s) \frac{\partial Q_v}{\partial P_\alpha}}$$

has lead characteristics and a static gain of less than unity, permitting higher loop gain, faster flow-control response, and much higher flow controller gain. To avoid problems with flow controller tuning, the system should always be operated in one flow regime or the other. In most cases steam flow is noncritical.

Signal Flow Diagram Simplification

Since we have called for averaging level control, the natural frequency of the level control loop will be much lower than that of the flow control loop. Note that:

$$\begin{aligned} & K_v G_v(s) K_{cf} G_{cf}(s) \frac{\partial Q_v}{\partial X_v} B(s) \\ &= \frac{K_v G_v(s) K_{cf} G_{cf}(s) \frac{\partial Q_v}{\partial X_v} \left(\frac{1}{1 - \psi(s) \frac{\partial Q_v}{\partial P_\alpha}} \right) K_{mf} G_{mf}(s)}{1 + K_v G_v(s) K_{cf} G_{cf}(s) \frac{\partial Q_v}{\partial X_v} \left(\frac{1}{1 - \psi(s) \frac{\partial Q_v}{\partial P_\alpha}} \right) K_{mf} G_{mf}(s)} \\ &\quad \times \frac{1}{K_{mf} G_{mf}(s)} \approx \frac{1}{K_{mf}} \end{aligned} \quad (15.44)$$

This leads us to the final signal flow diagram of Figure 15.11.

Base Level Control by Direct Manipulation of Steam Valve

From Figure 15.9 and by assuming $\Delta(s) = 0$, we can prepare the signal flow diagram of Figure 15.12. Note that if steam flow is critical, $\partial Q_v / \partial P_\alpha = 0$.

Further Mathematical Simplification

It has already been indicated that for averaging level control cascaded to steam flow control, one may substitute $1/K_{mf}$ for the flow control loop. Use of the mathematical models discussed here on commercial reboilers indicates that typical time constants range from a fraction of a second to 5–10 seconds. Practically speaking $\gamma(s)$, $\phi(s)$, and $\Delta(s)$ reduce to constants. As we will see in Chapters 16 and 17, it usually will be possible to use much simpler reboiler models than that discussed in this section.

15.4 PARTIALLY FLOODED REBOILERS

The partially flooded reboiler is similar in many ways to the partially flooded condenser. Usually, although not always, it is a vertical thermosyphon reboiler. As discussed in Chapter 4, Section 2, it is controlled by throttling the steam condensate, which in turn varies the condensate level in the shell and thereby the heat-transfer area for condensation. That area covered by liquid permits only sensible heat transfer from the condensate; this is a small heat load compared with that of the condensing steam and is treated as negligible.

The signal flow diagram of Figure 15.5 for a flooded condenser may be used as a starting point. In this case pressure dynamics are essentially negligible. If the steam supply pressure is constant, the steam condensing temperature is also constant.

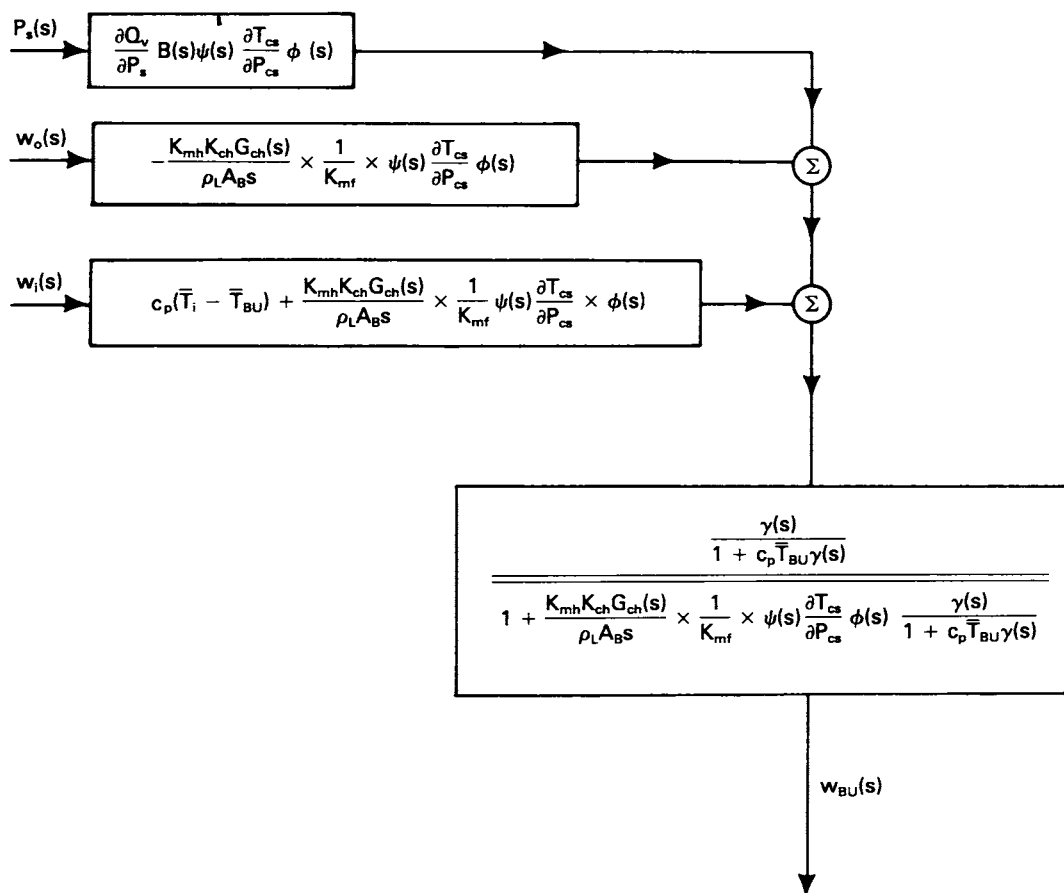


FIGURE 15.11

Final signal flow diagram for base level control cascaded to steam flow control

The signal flow diagram of Figure 15.13 may be prepared next. As shown by the reduced form of Figure 15.14:

$$\frac{w_c(s)}{w_o(s)} = \frac{1}{\frac{\lambda_{st}}{U_R \bar{\Delta T} \frac{\partial A_R}{\partial W_{SH}}} s + 1}$$

(15.45)

where

- w_c

=

rate of steam condensation, lbm/min
- w_o

=

rate of steam condensate withdrawal, lbm/min
- λ_{st}

=

steam latent heat, pcu/lbm
- U_R

=

heat-transfer coefficient, $\frac{\text{pcu}/^\circ\text{C}}{\text{ft}^2 \text{ min}}$
- \bar{A}_R

=

average exposed heat-transfer area, ft², for steam condensate
- $\frac{\partial A_R}{\partial W_{SH}}$

=

$\frac{A_{RT}}{[W_{SH}]_{\max}}$
- A_{RT}

=

total heat-transfer area of reboiler, ft²
- W_{SH}

=

condensate on shell side, lbm

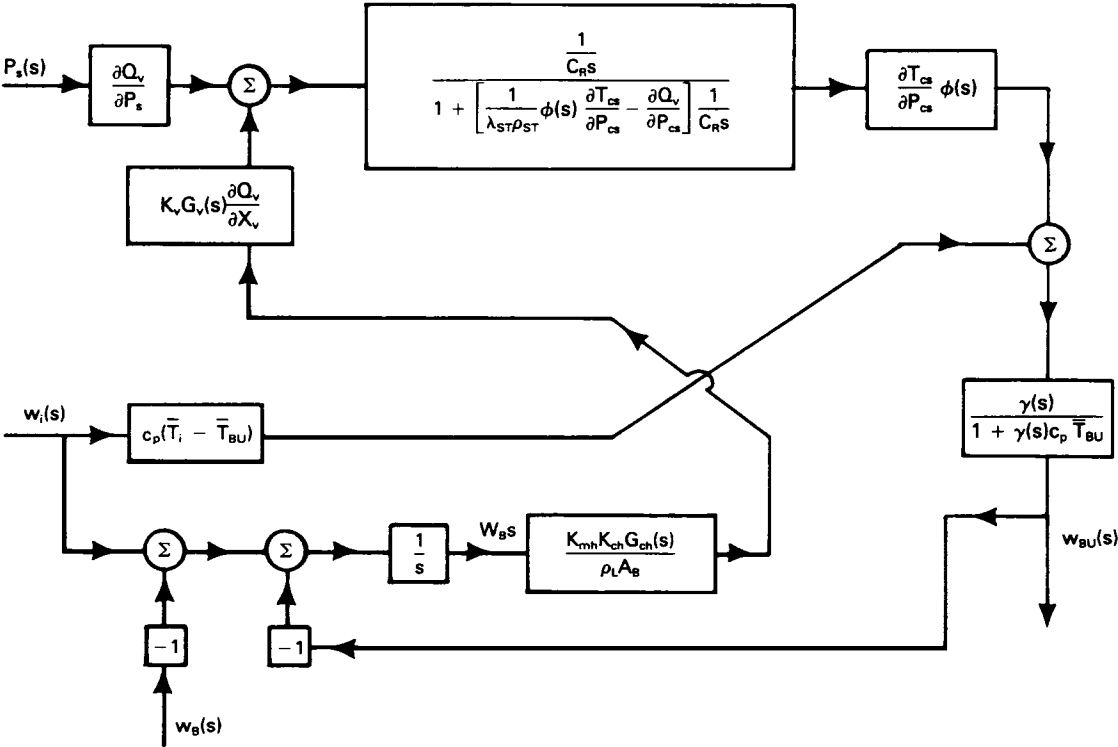


FIGURE 15.12
Signal flow diagram for base level control by
direct manipulation of steam valve

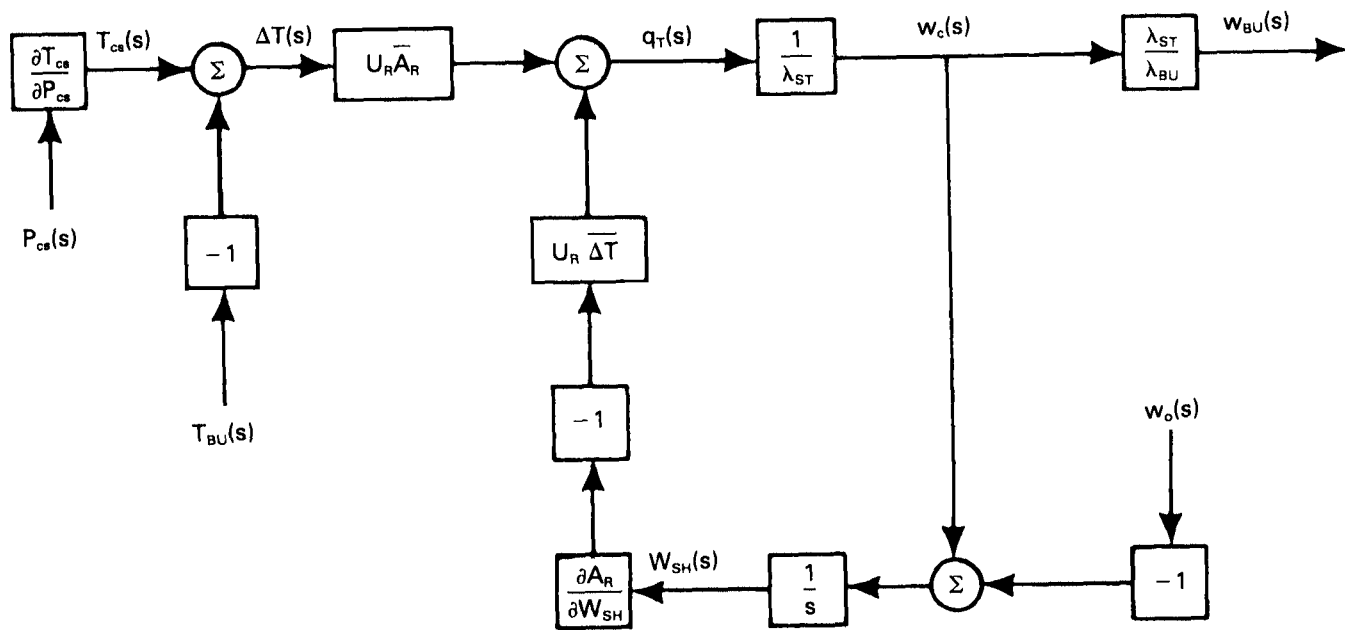


FIGURE 15.13
Preliminary signal flow diagram for flooded reboiler

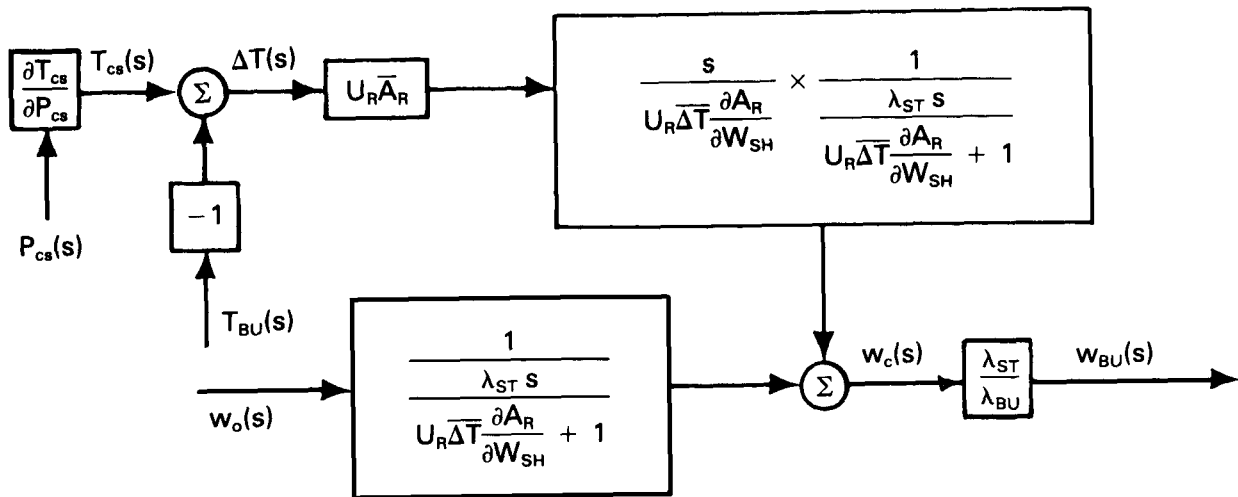


FIGURE 15.14
Reduced signal flow diagram for flooded reboiler

$[W_{SH}]_{\max}$ = weight of condensate that will fill shell side of reboiler, lbm
 w_{BU} = boilup, lbm/min
 λ_{BU} = latent heat of boiling liquid, pcu/lbm
 ΔT = $\bar{T}_{ST} - \bar{T}_{BU}$, °C

Design experience with flooded reboilers is limited but indicates that typical time constants are of the order of 2–5 minutes. Simulation studies show that substantial improvement in response speed may be achieved by lead-lag compensations with transfer functions such as:

$$\frac{\tau_D s + 1}{\frac{\tau_D}{\alpha} s + 1}$$

where we let:

$$\tau_D = \frac{\lambda_{st}}{U_R \bar{\Delta T} \frac{\partial A_R}{\partial W_{SH}}} = \tau_{RB} \quad (15.46)$$

Commercial lead-lag compensators commonly have values of α between 6 and 30. Some provide a fixed α and some have adjustable α . The former are usually much less expensive.

It is interesting to look at the response of w_c to a change in ΔT . From Figure 15.14:

$$\frac{w_c(s)}{\Delta T(s)} = U_R \bar{A}_R \left[\frac{\tau_{RB} s}{\lambda_{st}} \times \frac{1}{\tau_{RB} s + 1} \right] \quad (15.47)$$

If w_o were held constant, we would intuitively expect a step increase in ΔT to cause an initial increase in heat transfer, and therefore in w_c . Condensate level, however, would eventually increase (thereby decreasing A_R) and eventually w_c would have to equal w_o . Solving equation (15.47) for a step increase in ΔT shows this to be true:

$$\Delta w_c(t) = \mathcal{L}^{-1} \left[\frac{\Delta(\Delta T)}{s} U_R \bar{A}_R \left[\frac{s}{\lambda_{st}} \times \frac{1}{s + \frac{1}{\tau_{RB}}} \right] \right] \quad (15.48)$$

so that:

$$\Delta w_c(t) = \Delta(\Delta T) \left(\frac{U_R \bar{A}_R}{\lambda_{st}} \right) e^{-t/\tau_{RB}} \quad (15.49)$$

In equation (15.47) $w_c(t)$ is a perturbation variable (deviation from steady state). As shown by equation (15.49), it decays to zero as time goes to infinity.

15.5 PARTIALLY FLOODED REBOILERS FOR LOW-BOILING MATERIALS

In Chapter 4, Section 2, we discussed a variation in flooded reboiler design for low-boiling materials. As shown by Figure 4.4, we throttle the steam instead of the condensate. The static force–balance relationship is given by equation (4.1):

$$H_L \rho_L \frac{g_L}{g_c} = H_S \rho_L \frac{g_L}{g_c} + P_\sigma + \Delta P_{\text{line}} \quad (4.1)$$

where

H_L = loop seal or standpipe height, feet

H_S = condensate level in the shell, feet

P_σ = steam pressure in shell, lbf/ft² abs

ρ_L = condensate density, lbm/ft³

The vented loop seal maintains P_σ just a little above atmospheric pressure.

We can now write the following equations:

$$P_\sigma(s) = \frac{w_s(s) - w_c(s)}{s \rho_v (\bar{V}/\bar{P}_\sigma)} \quad (15.50)$$

$$T_\sigma(s) = \frac{\partial T_\sigma}{\partial P_\sigma} P_\sigma(s) \quad (15.51)$$

From equation (4.1), if we assume that the line loss ΔP_{line} is negligible, and that $H_L \rho_L \frac{g_L}{g_c}$ is constant, then:

$$H_s(s) = \frac{-P_\sigma(s)}{\rho_L (g_L/g_c)} \quad (15.52)$$

$$\frac{\partial A_R}{\partial H_s} = \frac{A_{RT}}{[H_s]_{\text{max}}} \quad (15.53)$$

$$A_R(s) = \frac{\partial A_R}{\partial H_s} H_s(s) \quad (15.54)$$

$$q_T(t) = U_R A_R(t) \Delta T(t) \quad (15.55)$$

$$\Delta T = T_\sigma - T_{BU} \quad (15.56)$$

$$w_c = \frac{q_T}{\lambda_\sigma} \quad (15.57)$$

$$w_{BU} = \frac{q_T}{\lambda_{BU}} \quad (15.58)$$

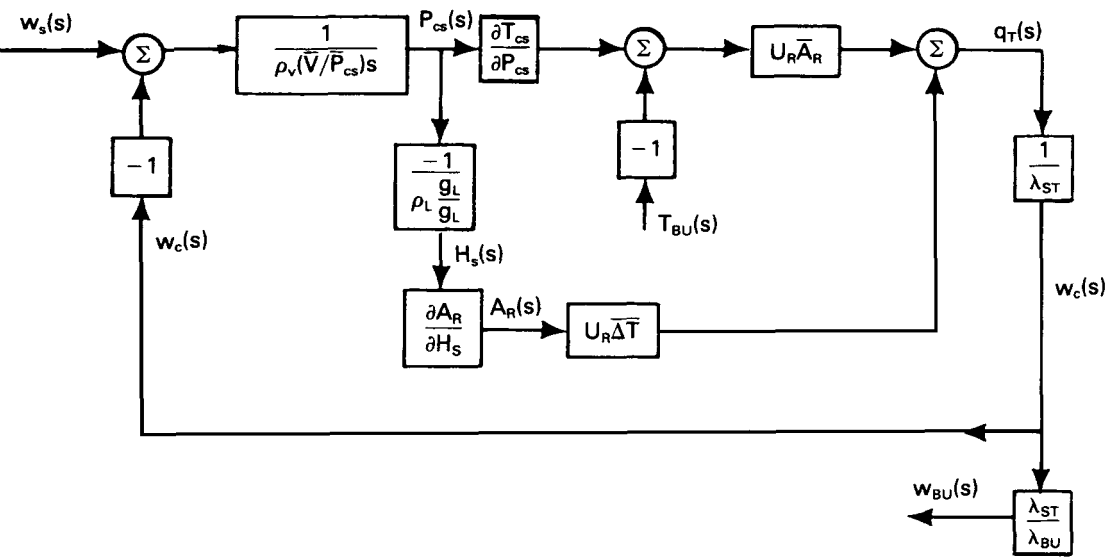


FIGURE 15.15
Signal flow diagram for flooded reboiler
for low boiling point materials

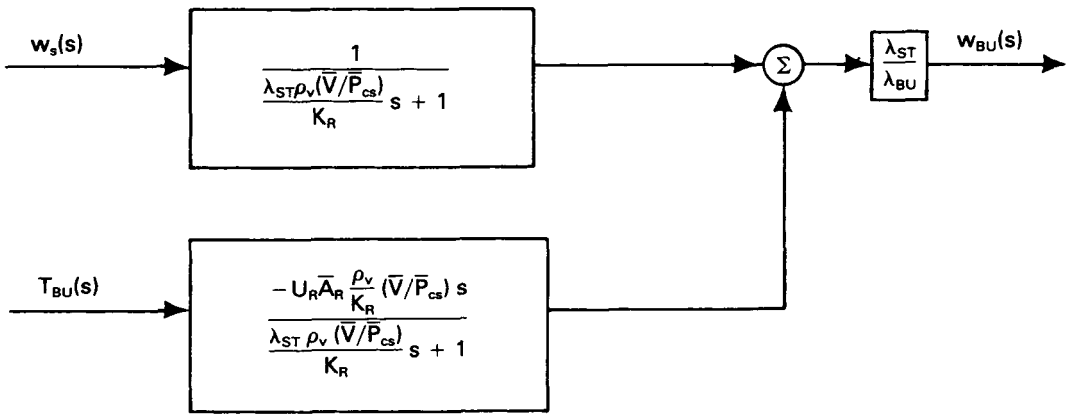


FIGURE 15.16
Reduced signal flow diagram
for flooded reboiler for low boiling point materials

where

w_s = steam flow, lbm/min

w_c = rate of steam condensation, lbm/min

ρ_v = steam density in shell, lbm/ft³

\bar{V} = average free volume in shell above liquid level, ft³

T_α = condensing temperature, °C

ρ_L = condensate density, lbm/ft³

Other terms are as defined earlier in this chapter.

To get equation (15.55) into the s domain, write:

$$q_T(s) = \frac{\partial q_T}{\partial A_R} A_R(s) + \frac{\partial q_T}{\partial \Delta T} \Delta T(s) \quad (15.59)$$

where

$$\frac{\partial q_T}{\partial A_R} = U_R \bar{\Delta T} \quad (15.60)$$

$$\frac{\partial q_T}{\partial \Delta T} = U_R \bar{A}_R \quad (15.61)$$

We may now prepare the signal flow diagram of Figure 15.15, which may then be reduced to the form of Figure 15.16. From the latter we see that:

$$\frac{w_c(s)}{w_s(s)} = \frac{1}{\frac{\lambda_{st} \rho_v (\bar{V}/\bar{P}_\alpha)}{K_R} s + 1} \quad (15.62)$$

where

$$K_R = \frac{\partial T_\alpha}{\partial P_\alpha} U_R \bar{A}_R - \frac{\partial A_R / \partial H_s}{\rho_L \frac{g_L}{g_c}} U_R \bar{\Delta T} \quad (15.63)$$

Also:

$$\frac{w_c(s)}{T_{BU}(s)} = \frac{-U_R \bar{A}_R}{K_R} \rho_v (\bar{V}/\bar{P}_\alpha) \times \frac{s}{\tau_{RB} s + 1} \quad (15.64)$$

As in the previous section, if there were a step change in T_{BU} , there would be an immediate spike in w_c , which would slowly decay to zero.

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16

Liquid Level Control

16.1 INTRODUCTION

for distillation columns level control refers, in most cases, to overhead condenser receivers (reflux accumulators), column bases, surge or feed tanks, and sometimes steam condensate receivers. For most of these applications, the chief function is not that of holding level constant, but rather of achieving the smoothest possible transitions in manipulated flows in response to disturbances. This is “averaging” level control.¹

As stated in Chapter 1, the functions of averaging level control are:

1. Balancing inflows against outflows at a point in a process.
2. Providing for smooth and gradual changes in manipulated flows to avoid upsetting process equipment.
3. Maintaining inventory or accumulation between an upper and a lower limit (not at a fixed value).

On new projects the engineer is confronted with two alternatives: (1) the tank or holdup size is already specified and the problem is to get maximum flow smoothing, or (2) the tank is to be sized and the level control system designed to achieve flow smoothing adequate for downstream composition controls. For the latter we make $\tau_H \geq 10/\omega_R$, where ω_R is downstream closed-loop resonant frequency, radians/minute.

Before getting into specific applications of level control on distillation columns, let us review briefly the theory of averaging level control on simple vessels.

16.2 LEVEL CONTROL OF SIMPLE VESSELS

For a simple vessel such as shown in Figure 16.1 where the level is controlled by outflow and there is no significant level self-regulation effect, we need only a few equations:

$$\frac{Q_i(s) - Q_o(s)}{As} = H(s) \quad (16.1)$$

$$\theta_{mh}(s) = K_{mh} H(s) \quad (16.2)$$

$$\theta_c(s) = K_{ch} G_{ch}(s) \times \theta_{mh}(s) \quad (16.3)$$

$$Q_o(s) = \frac{dQ_o}{d\theta_c} \theta_c(s) \quad (16.4)$$

where

Q_i = inflow, ft³/min

Q_o = outflow, ft³/min

A = vessel cross-sectional area, ft² (vertical, cylindrical vessel assumed)

H = liquid level, feet

θ_{mh} = level transmitter output signal

K_{mh} = 12 psi/ ΔH_T for pneumatics

ΔH_T = level transmitter input span, feet of process fluid

$K_{ch} G_{ch}(s)$ = controller transfer function

θ_c = controller output signal

$dQ_o/d\theta_c$ = valve gain, or flow control loop gain, $\frac{\text{ft}^3/\text{min}}{\text{psi}}$ (see discussions in Sections 3.9 and 4.7)

For a cascade level-flow system:

$$\frac{dQ_o}{d\theta_c} = \frac{1}{K_{mf}}$$

where

K_{mf} = flow measurement gain of linear flow meter

$$= \left(\frac{12}{[Q_o]_m} \right)$$

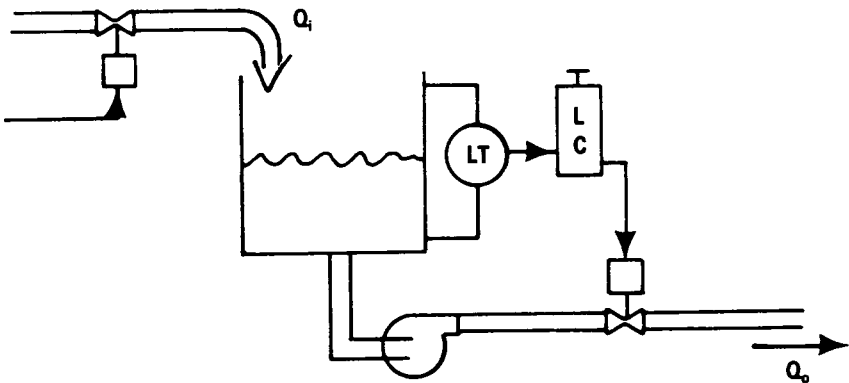


FIGURE 16.1
Level control of simple vessel

where

$[Q_o]_m$ = maximum flow of flow-meter span, ft^3/min

The analysis employed here is further simplified in that the effects of variable valve-pressure drop are omitted, as are transmitter dynamics.

Equations (16.1) through (16.4) may be combined into the signal flow diagram of Figure 16.2, from which we may write by inspection:

$$\frac{Q_o(s)}{Q_i(s)} = \frac{\frac{K_{mb}}{As} K_{ch} G_{ch}(s) \frac{dQ_o}{d\theta_c}}{1 + \frac{K_{mb}}{As} K_{ch} G_{ch}(s) \frac{dQ_o}{d\theta_c}} \quad (16.5)$$

and

$$\frac{H(s)}{Q_i(s)} = \frac{\frac{1}{As}}{1 + \frac{K_{mb}}{As} K_{ch} G_{ch}(s) \frac{dQ_o}{d\theta_c}} \quad (16.6)$$

Proportional-Only Control

For this case $K_{ch} G_{ch}(s)$ becomes simply K_{ch} . Then equation (16.5) becomes:

$$\frac{Q_o(s)}{Q_i(s)} = \frac{1}{\frac{A}{K_{mb} K_{ch} \frac{dQ_o}{d\theta_c}} s + 1} \quad (16.7)$$

and equation (16.6) becomes:

$$\frac{H(s)}{Q_i(s)} = \frac{1}{K_{mb} K_{ch} \frac{dQ_o}{d\theta_c}} \times \frac{1}{\frac{A}{K_{mb} K_{ch} \frac{dQ_o}{d\theta_c}} s + 1} \quad (16.8)$$

If the input span of the valve positioner is the same as the transmitter output span (as, for example, 3–15 psig), and if the valve has an installed linear flow characteristic with a wide-open capacity approximately equal to four times flowsheet flow, Q_{FS} , then:

$$\frac{A}{K_{mb} K_{ch} \frac{dQ_o}{d\theta_c}} = \frac{A}{\left[\frac{12}{\Delta H_T} \right] K_{ch} \left[\frac{4 Q_{FS}}{12} \right]} = \frac{A \Delta H_T}{K_{ch} \times 4 Q_{FS}} = \tau_H \quad (16.9)$$

This last equation is very useful for finding the desired holdup, $A \Delta H_T$, provided a value of τ_H is specified. Usually we choose $K_{ch} = 2$ and bias the

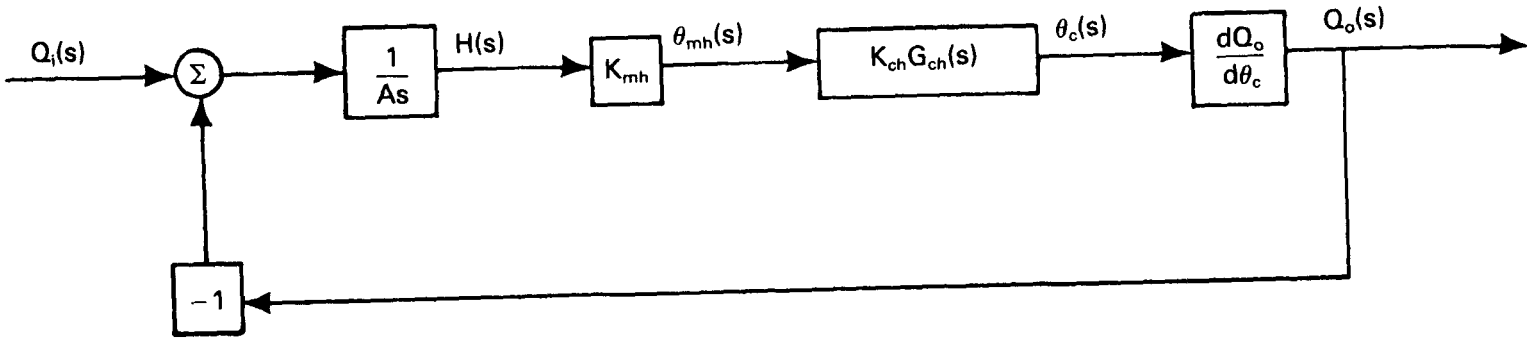


FIGURE 16.2
Signal flow diagram for simple level control system

proportional controller so that the tank level is midscale on the level transmitter. This leaves the top 25 percent of the transmitter span for overrides. Then:

$$A \Delta H_T = V_T = \tau_H K_{ch} (4 Q_{FS}) \quad (16.9a)$$

$$= 8 \tau_H Q_{FS} \quad (16.9b)$$

For level control cascaded to flow control:

$$A \Delta H_T = V_T = \tau_H K_{ch} (Q_o)_{\max} \quad (16.9c)$$

Proportional-Reset Control

Although the proportional-only control system is simple, inexpensive, and almost foolproof (at least when implemented with fixed-gain relays), it has limitations:

1. If tank size is specified, the flow smoothing is limited by τ_H because one cannot safely use $K_{ch} < 1$.
2. Because mechanics often do not calibrate valves precisely, use of $K_{ch} = 1$ is risky; the valve may not be closed when the level is at zero. Consequently we usually specify $K_{ch} = 2$.
3. If tank size is not specified but is to be calculated from equation (16.9a) or (16.9c), a large tank will be required since for a given τ_H , V_T is proportional to K_{ch} .

Although in theory there exists a number of controllers that permit one to use $K_{ch} < 1$, the PI controller has been most popular. Since the unenhanced PI controller with $K_{ch} < 1$ does not ensure that the tank will not run dry or overflow, past practice has been to have high and low alarms or high and low interlocks. In recent years, however, the PI controller enhanced or augmented with auto overrides has provided an almost foolproof way of keeping liquid within the vessel. It provides, under most circumstances, much more flow smoothing for a given size vessel than will a proportional-only controller.¹⁰

Before exploring the theory, let us make some additional design assumptions:

1. Nozzle-to-nozzle spacing is so chosen that process operation will be satisfactory with the level at any location between the nozzles. "Nozzles" here refers to those used for connecting the level-measuring device to the vessel.
2. Normal set point for the PI level controller is midscale of level transmitter span, that is, $\Delta H_T/2$. This is required for proper functioning of the auto overrides. Note that the level transmitter span is usually less than the nozzle-to-nozzle spacing. This allows for some variation in liquid specific gravity.
3. Level control is cascaded to flow control. With floating pressure columns and with the trend toward small control-valve pressure drops for energy conservation, this is virtually mandatory to counteract the effect of control valve up- and downstream pressure variations. The flow measurement must be linear; if an orifice flow meter is used, it must be followed by a square-root extractor.

A schematic for PI level control on a simple tank is given in Figure 16.3. As a result of a number of studies (unpublished), we have concluded that auto overrides with gain 2 and a controller tuned for a damping ratio of one are optimum for most situations. We have also found for most cases that the PI level control system so designed functions in a linear manner for step changes in input flow of up to 10 percent of span of the manipulated flow if $K_{ch} \geq 0.25$.¹⁰ This means that the controller output at its maximum value verges on being taken over by an auto override.

Other overrides are shown in both the level controller output signal path to the flow controller set point and in the flow controller output signal path to the control valve. The latter arrangement (overrides in the signal path to the control valve) has been far more common, but the former permits more accurate, quantitative design. The former also implies that whenever the flow control station is not switched to "remote auto," the overrides are out of service. This may or may not be desirable. But regardless of override location (other than auto overrides), we provide a switching design that causes the level controller reset to be bypassed (i.e., level controller has very fast reset) whenever the flow control station is not switched to "remote auto." This virtually eliminates "bumping" when the flow controller switches from either "manual" or "local auto" to "remote auto." A primary control station is not necessary, but a level indicator is desirable.

In the analysis that follows we will ignore the role of overrides and will assume that the PI level controller is always "in command." This assumption permits us to use Laplace transforms and frequency response. If it is desired to predict system behavior when forced by disturbances large enough to cause an override to take over, we must resort to digital simulation.

The transfer function for a PI controller is:

$$K_{ch} G_{ch}(s) = K_{ch} \left(\frac{\tau_R s + 1}{\tau_R s} \right)$$

where τ_R is the reset time in minutes. Substituting this into equation (16.5) and replacing dQ_o/dQ_c by $1/K_{mf}$, we get:

$$\frac{Q_o(s)}{Q_i(s)} = \frac{\frac{K_{mh}}{A_s} K_{ch} \left(\frac{\tau_R s + 1}{\tau_R s} \right) \frac{1}{K_{mf}}}{1 + \frac{K_{mh}}{A_s} K_{ch} \left(\frac{\tau_R s + 1}{\tau_R s} \right) \frac{1}{K_{mf}}} \quad (16.10)$$

$$= \frac{\tau_R s + 1}{\frac{A \tau_R K_{mf}}{K_{mh} K_{ch}} s^2 + \tau_R s + 1} \quad (16.11)$$

$$= \frac{\tau_R s + 1}{\tau_H \tau_R s^2 + \tau_R s + 1} \quad (16.12)$$

Similarly, from equation (16.6):

$$\frac{H(s)}{Q_i(s)} = \frac{\frac{\tau_H \tau_R}{A} s}{\tau_H \tau_R s^2 + \tau_R s + 1} \quad (16.13)$$

Now the denominator of equations (16.12) and (16.13) has some interesting characteristics not widely appreciated. It is a quadratic whose damping ratio is:

$$\zeta = \frac{\tau_R}{2\sqrt{\tau_R \tau_H}} = \frac{1}{2} \sqrt{\frac{\tau_R}{\tau_H}} = \frac{1}{2} \sqrt{\tau_R \left(\frac{K_{mh} K_{ch}}{A K_{mf}} \right)} \quad (16.14)$$

or

$$\tau_R = 4 \zeta^2 \tau_H \quad (16.14a)$$

Let us also define:

$$\tau_Q = \sqrt{\tau_R \tau_H} = \sqrt{\tau_R \frac{A K_{mf}}{K_{mh} K_{ch}}} \quad (16.15)$$

$$= 2 \zeta \tau_H \quad (16.15a)$$

From equations (16.14) and (16.15) we can see that if τ_R is fixed, then as one decreases K_{ch} two things happen:

1. The damping ratio, ζ , approaches zero and the control loop becomes very resonant, approaching instability.
2. τ_Q becomes very large.

The loop therefore becomes slower and less stable at the same time. This resonance is sometimes called a “reset cycle” since it would not exist if the controller did not have automatic reset.

If, on the other hand, one increases K_{ch} while holding τ_R constant, τ_Q becomes small, transmitter and valve dynamics become significant, and the loop eventually becomes unstable. This is commonly called a “gain cycle” since it is caused by excess gain. Since the loop approaches instability for both very large and very small values of K_{ch} , we say that it is *conditionally* stable.

In designing a level control system with a proportional-reset controller several practical considerations must be kept in mind:

1. The damping ratio preferably should be at least unity. A low damping ratio, as shown by equations (16.12) and (16.13), causes severe peaking in the frequency response in the vicinity of the closed-loop natural frequency, $1/\tau_Q$. Flow and level regulation in that frequency range will be very poor.
2. Adjacent or related process controls must be designed with closed-loop natural frequencies much different from that of the level control; usually they are designed to be much faster.

3. For a given damping ratio, reset time must be increased as K_{cb} is decreased. It frequently happens that for a desired τ_Q and specified K_{cb} , one cannot readily obtain the necessary τ_R with a particular commercial controller. The major instrument manufacturers can usually furnish modification kits or modified controllers with a larger τ_R .

For $\zeta = 1$ it can be seen from equation (16.14a) that:

$$\tau_R = 4 \tau_H \quad (16.16)$$

and

$$\tau_Q = 2 \tau_H = \tau_R/2 \quad (16.17)$$

For $\zeta = 2$,

$$\tau_R = 16 \tau_H \quad (16.18)$$

and

$$\tau_Q = 4 \tau_H = \tau_R/4 \quad (16.19)$$

One of the major disadvantages of PI controllers for liquid level is that they always cause the manipulated flow change temporarily to be greater than the disturbance flow change. For example, if the system we have been considering is subjected to a step change in inflow, we obtain the following:

$$\begin{aligned} \frac{(\Delta Q_o)_{\max}}{\Delta Q_i} &= 1.38 \quad \text{for } \zeta = 0.4 \\ &= 1.14 \quad \text{for } \zeta = 1.0 \\ &= 1.048 \quad \text{for } \zeta = 2.0 \end{aligned}$$

The fact that the outflow swings more than the inflow can create serious problems if the process is running close to capacity. For such applications we should choose $\zeta = 2.0$ or use a proportional-only controller. For most other applications, $\zeta = 1$ should suffice and places less of a burden on available controller settings.

At this point it may be appropriate to note that a viable alternative to PI level control is PL level control.^{6,7} It has transfer functions very similar to those of the PI level control; for a damping ratio of unity, the transfer functions are identical if one reduces the PL level controller gain, K_{cb} , by a factor of 2. For $K_{cb} < 1$, it requires auto overrides just as the PI controller does. It has the feature, useful in some circumstances, of not needing antireset windup. It is not a standard commercial item but usually can be assembled with various standard devices.

Augmented PI Controllers

A plain PI controller, even if tuned for $\zeta = 1$, cannot guarantee that level will be held within the vessel. To protect upper and lower permissible level limits, we have found two approaches useful:

1. Auto overrides for pneumatics. When the level becomes too high or too low, a proportional-only controller (usually a fixed-gain relay) takes over through a high- or low-selector.² As shown by Figure 16.3, the high-level gain 2 auto override is so biased that its output is 3.0 psig when the input (level transmitter signal) is 9.0 psig. Then its output is 15.0 psig when its input is 15 psig. Correspondingly the low-level gain 2 auto override is so biased that when the input is 9.0 psig, its output is 15 psig; if its input goes down to 3.0 psig (zero level), its output is also 3.0 psig.

2. Nonlinear PI controllers for electronics.⁴ A preferred version has long reset time and a small K_{θ} in the vicinity of the set point. As the level deviates significantly from set point, K_{θ} increases and τ_R decreases. Although in theory this is not quite as foolproof as auto overrides, our studies show that it rarely permits excessive deviations of level, and then only by a small amount. This design does not provide flow smoothing quite as good as that of the PI plus override scheme. This is so because manipulated flow changes more rapidly so that level moves away from the top or bottom of the tank more quickly.

Effect of Installed Valve Flow Characteristic

If the PI level control system (or proportional-only) is not a cascade level-flow system, then it is desirable to have a control valve with a linear installed flow characteristic, that is, $dQ_o/d\theta_c = \text{constant}$. Then control-loop dynamics would be independent of flow rate and

$$\frac{dQ_o}{d\theta_c} = \frac{(Q_v)_{\max}}{12} \frac{\text{ft}^3/\text{min}}{\text{psi}} \quad (\text{or gpm/psi})$$

where $(Q_v)_{\max}$ is the flow through the valve in its wide-open position.

If the valve has an equal-percentage installed characteristic, then from Table 15.1, reference 1:

$$\frac{dQ_o}{d\theta_c} = \frac{\bar{Q}_o k_{EP}}{12}$$

Since $k_{EP} = 3.9$ for a 50:1 equal-percentage valve ($k_{EP} = \ln \alpha$ where $\alpha = 50$):

$$\frac{dQ_o}{d\theta_c} \approx \frac{4 \bar{Q}_o}{12} = \frac{\bar{Q}_o}{3}$$

Referring to equation (16.14), we see that if a controller is set up correctly with $\zeta = 1$ at $Q = Q_{FS}$, then:

- $\zeta = 0.7$ when $Q_o = (Q_{FS}/2)$, and
- $\zeta = 1.4$ when $Q_o = 2 Q_{FS}$

Thus we see that, with an equal-percentage installed flow characteristic, the relative stability is decreased at low flow and increased at high flow. For this case it is best to find controller settings for $\zeta = 1$ at the minimum expected flow.

If pump and valve curves are available, and if the hydraulic resistance and static heads of the process equipment are known, we can usually calculate the installed flow characteristic. Alternatively, with permission from production supervision, we can experimentally make a plot of valve loading signal versus flow. The slope of this curve at average rate \bar{Q}_o is $dQ_o/d\theta_v$.

For a cascade level-flow system, the control-valve installed-flow characteristic should also be linear.⁵

Auto-Override Time Constant

For large swings in the disturbance flow that will drive the level far enough that one of the auto overrides takes effect, the system temporarily will be under control of the proportional-only auto override. In this regime it is not absolutely necessary that control be stable; the auto override will drive the system back into the linear, stable regime. The proportional-only auto-override control system has the characteristic time constant:

$$[\tau_H]_{OR} = \frac{A \Delta H_T}{K_{OR} (Q_o)_{\max}}$$

To date we have been unable to come up with a truly rational way of specifying the desired numerical value of $[\tau_H]_{OR}$. On the basis of experience, we suggest that it not be less than one minute. For some applications a much larger value will be desirable.

Tuning Procedure for PI Controller

From the preceding it may be apparent that to get maximum flow smoothing with a given commercial controller and specified damping ratio, the engineer will be constrained by either $[K_{cb}]_{\min}$ or $[\tau_R]_{\max}$ of that controller. To find out which, and to determine $[K_{cb}]_{\text{design}}$ or $[\tau_R]_{\text{design}}$, the following procedure is suggested:

1. Let $[K_{cb}]_{\min} = 0.25$ (see reference 10). This is larger than the minimum available gain of any commercial controller with which we are familiar.
2. If

$$\frac{A \Delta H_T}{(Q_o)_m} > \frac{[\tau_R]_{\max} [K_{cb}]_{\min}}{4 \zeta^2}$$

choose

$$[\tau_R]_{\text{design}} = [\tau_R]_{\max}$$

and find

$$[K_{cb}]_{\text{design}} = \frac{4 \zeta^2}{[\tau_R]_{\max}} \times \frac{A \Delta H_T}{[Q_o]_m}$$

3. If

$$\frac{A \Delta H_T}{(Q_o)_m} < \frac{[\tau_R]_{\max} [K_{ch}]_{\min}}{4 \zeta^2}$$

choose

$$[K_{ch}]_{\text{design}} = [K_{ch}]_{\min}$$

and find

$$[\tau_R]_{\text{design}} = \frac{4 \zeta^2}{[K_{ch}]_{\min}} \times \frac{A \Delta H_T}{[Q_o]_m}$$

4. If

$$\frac{A \Delta H_T}{(Q_o)_m} = \frac{[\tau_R]_{\max} [K_{ch}]_{\min}}{4 \zeta^2}$$

$$[K_{ch}]_{\text{design}} = [K_{ch}]_{\min}$$

and

$$[\tau_R]_{\text{design}} = [\tau_R]_{\max}$$

16.3 LEVEL CONTROL OF OVERHEAD CONDENSER RECEIVER VIA TOP-PRODUCT WITHDRAWAL

Ordinarily this is a simple system to design since it almost always fits the preceding analysis. Chapter 3, Section 9, discusses some of the practical details. For proportional-only control, one should make $\tau_H \geq 2$ minutes; for a PI controller, one should make $\tau_Q \geq 2$ minutes.

16.4 LEVEL CONTROL OF OVERHEAD CONDENSER RECEIVER VIA REFLUX MANIPULATION

For this application it is necessary to take reflux subcooling into account. It is also more convenient to use weight units. The amount of vapor condensed in the column by cold reflux is:

$$w_R \left[\frac{c_p}{\lambda} (T_o - T_R) \right]$$

where

w_R = external reflux flow, lbm/min

c_p = reflux specific heat, pcu/lbm °C

T_o = vapor condensing temperature, °C

T_R = external reflux temperature, °C

λ = vapor latent heat of condensation, pcu/lbm, of process vapor

Then

$$w_c(s) = w_{i-1}(s) - w_R(s) \left[\frac{c_p}{\lambda} (\bar{T}_o - \bar{T}_R) \right] \quad (16.20)$$

$$W_T(s) = \frac{w_c(s) - w_D(s) - w_R(s)}{s} \quad (16.21)$$

$$H_T(s) = \frac{W_T(s)}{\rho_L A_T} \quad (16.22)$$

where

w_{i-1} = vapor *to* top tray, lbm/min

w_c = vapor *from* top tray to condenser, lbm/min; also rate of condensation

w_D = top-product flow, lbm/min

ρ_L = external reflux density, lbm/ft³

A_T = cross-sectional area of tank, ft² (vertical cylindrical design assumed)

Strictly speaking we should include condenser dynamics but these usually amount to only a half-minute to a minute lag. The larger τ_H or τ_Q is, the less significant will be condenser lag. By making $\tau_H \geq 5$ minutes, we can usually ignore condenser lag.

We may now prepare the signal flow diagram of Figure 16.4 for proportional-only level control. By inspection we can see that:

$$\begin{aligned} \frac{w_R(s)}{w_{i-1}(s) - w_D(s)} &= \frac{1}{\left[1 + \frac{c_p}{\lambda} (\bar{T}_o - \bar{T}_R) \right]} \\ &\times \frac{1}{\frac{\rho_L A_T s}{K_{mb} K_{cb} \frac{dw_R}{d\theta_c} \left[1 + \frac{c_p}{\lambda} (\bar{T}_o - \bar{T}_R) \right]} + 1} \end{aligned} \quad (16.23)$$

Let us now define:

$$K_{SC} = 1 + \frac{c_p}{\lambda} (\bar{T}_o - \bar{T}_R) \quad (16.24)$$

Then

$$\frac{w_R(s)}{w_{i-1}(s) - w_D(s)} = \frac{1}{K_{SC}} \times \frac{1}{\tau_H s + 1} \quad (16.25)$$

where

$$\tau_H = \frac{\rho_L A_T}{K_{SC} K_{mb} K_{cb} \frac{dw_R}{d\theta_c}}$$

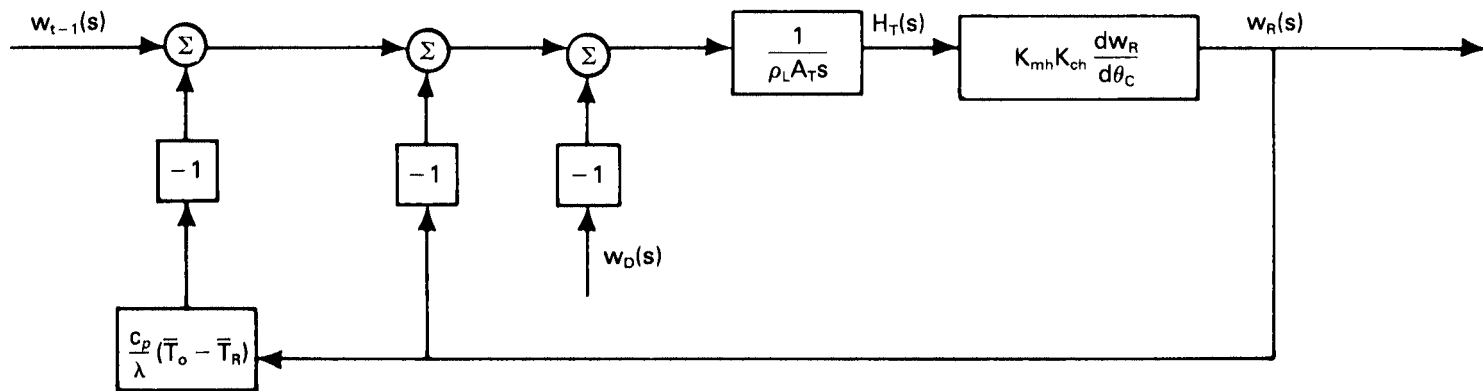


FIGURE 16.4

Signal flow diagram for proportional-only condensate seal pot level control via reflux flow manipulation

Note that τ_H contains K_{SC} and is inversely proportional to it. Some idea of the magnitude of K_{SC} may be obtained by assuming as an example $\bar{T}_O - \bar{T}_R = 100^\circ\text{C}$, $\lambda = 100$ pcu/lb, and $c_p = 0.5$ pcu/lb $^\circ\text{C}$. Then $K_{SC} = 1.5$. Subcooling therefore can have a significant effect on T_H .

For PI level control we can readily show that:

$$\frac{w_R(s)}{w_{t-1} - w_D(s)} = \frac{1}{K_{SC}} \times \frac{\tau_R s + 1}{\tau_H \tau_R s^2 + \tau_R s + 1} \quad (16.26)$$

16.5 COLUMN-BASE LEVEL CONTROL VIA BOTTOM-PRODUCT MANIPULATION

This is usually a fairly straightforward system to design since it fits the analysis of Section 16.2. Some of the practical details are discussed in Chapter 4. For most applications one should make $\tau_H \geq 10$ minutes.

16.6 COLUMN-BASE LEVEL CONTROL VIA FEED FLOW MANIPULATION

For this case the simple analysis of Section 16.2 must be extended to take into account the lags between the column feed point and the column base. Let us suppose that there are n trays between these two points and that hydraulic lag of each tray is first order with a time constant τ_{TR} (see Chapter 13). Typical values are in the range of 3–8 seconds. It has been shown elsewhere (Chapter 12 of reference 1) that a number of equal lags may be approximated by dead time:

$$\frac{1}{(\tau_{TR} s + 1)^n} \cong e^{-n \tau_{TR} s} \quad (16.27)$$

Such an approximation simplifies considerably either hand or computer calculations. If the feed enters at its boiling point, then

$$w_1(s) \cong e^{-n \tau_{TR} s} w_F(s) \quad (16.28)$$

and

$$W_B(s) = \frac{w_1(s) - w_v(s) - w_B(s)}{s} \quad (16.29)$$

where

w_1 = flow, lbm/min, from lowest downcomer into column base; flow due to reflux or feed change

w_v = boilup, lbm/min

w_B = bottom product, lbm/min

The equations for the remainder of the system now follow the analysis of Section 16.2. The complete system, level control cascaded to feed flow control, may be represented by the signal flow diagram of Figure 16.5. By inspection we can see that:

$$\frac{w_F(s)}{-[w_v(s) + w_B(s)]} = \frac{\frac{-K_{mb} K_{ch} G_{ch}(s)/K_{mfF}}{\rho_L A_B s}}{1 + \frac{K_{mb} K_{ch} G_{ch}(s)e^{-n\tau_{TR}s}/K_{mfF}}{\rho_L A_B s}} \quad (16.30)$$

Let us next define:

$$a = n \tau_{TR} \quad (16.31)$$

Without going through the mathematics, we will simply state that, by means of Bode and Nichols plots, we found that for well-damped response ($M_p = 2$ db):

$$\frac{a}{0.45} \leq \tau_H = \frac{\rho_L A_B K_{mfF}}{K_{mb} K_{ch}} = \frac{1}{K_L} \quad (16.32)$$

and

$$u_{res} = 0.25 = a \omega_R \quad (16.32a)$$

where

u_{res} = closed-loop resonant frequency, dimensionless

(ω_R) = closed-loop resonant frequency, rad/min

Equation (16.32) defines a minimum value of τ_H ; for greater stability and filtering one may use larger values of τ_H .

For PI control choose:

$$\tau_R \geq 5 \times \frac{1}{(0.25/a)} = 20 a \quad (16.33)$$

The level and inflow responses to a step change in outflow of an averaging level control system with dead time and a PI controller tuned as recommended above are given in Figure 16.6.

Recently we have done some work (unpublished) that suggests that incorporation of a Smith predictor^{8,9} would help greatly.

16.7 COLUMN-BASE LEVEL CONTROL CASCADED TO STEAM FLOW CONTROL

Ordinarily it is satisfactory to use liquid level as a measure of inventory. In the case of distillation columns, however, this measurement is ambiguous to some extent. The levels in the overhead condensate receiver and in the column

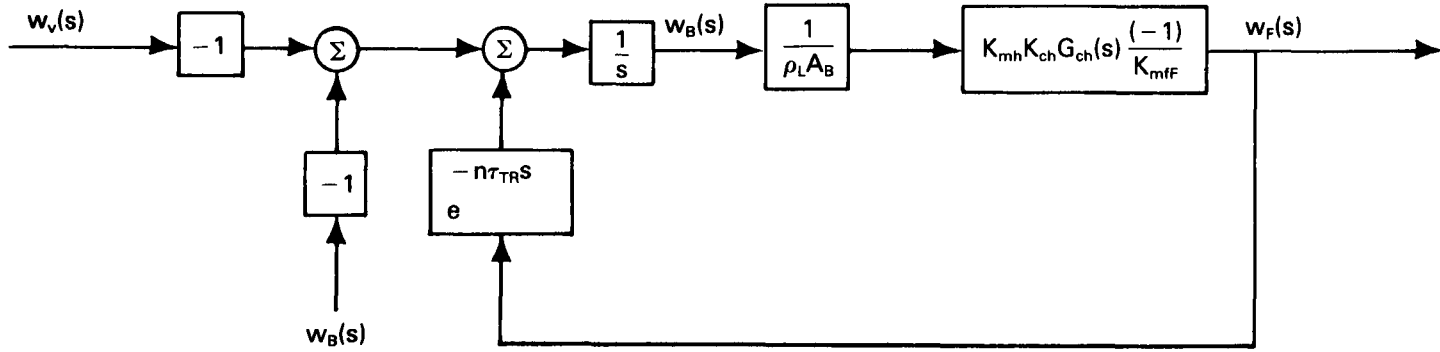


FIGURE 16.5
Signal flow diagram for proportional-only column base level control via feed flow manipulation

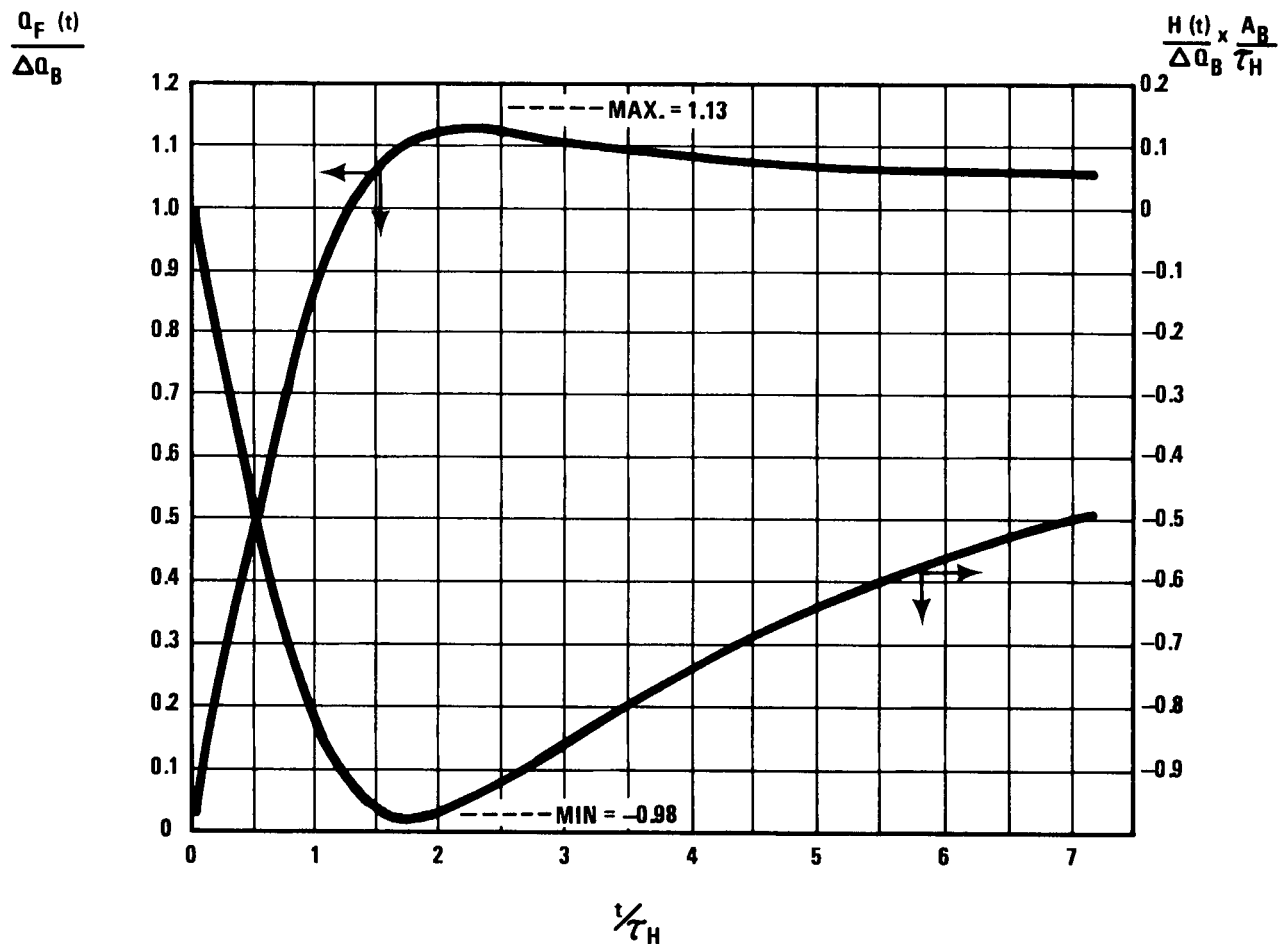


FIGURE 16.6
Responses of PI averaging level control system with dead time to step change in outflow

base do not accurately reflect either total column inventory or inventory changes. Put another way, level changes can occur that do not result from inventory changes. So far we have ignored these, but they can be very important when base level is controlled by throttling heating-medium flow.

Thermosyphon Reboiler Swell

In Chapter 4 it was mentioned that two factors can affect the volumetric percent vapor in the tubes and thereby cause a liquid displacement or “swell” in the column base.

A. Heat Load

For a constant liquid level in the column base, as heat load increases, percent vapor in the tubes increases. The effect is most marked at low heat loads. At a given operating point:

$$V_{HL}(s) = k_{v1} w_{ST}(s) \quad (16.34)$$

where

$$\begin{aligned} V_{HL} &= \text{change in tube vapor volume in tubes, ft}^3, \text{ due to heat load change} \\ &= \text{change in liquid volume in column base, ft}^3, \text{ due to heat load change} \\ w_{ST} &= \text{steam flow, lbm/min} \\ k_{v1} &= \frac{\partial V_{HL}}{\partial w_{ST}} \end{aligned}$$

B. Column-Base Liquid Level

For a constant heat load, an increase in base level normally decreases tube vapor volume:

$$V_H(s) = k_{v2} H_B(s) \quad (16.35)$$

where

$$\begin{aligned} V_H &= \text{change in tube vapor volume, ft}^3, \\ &\quad \text{due to base liquid level change} \\ &= \text{change in liquid volume in column base, ft}^3, \\ &\quad \text{due to change in tube vapor volume} \\ k_{v2} &= \frac{\partial V_H}{\partial H_B} \end{aligned}$$

The total volume change, $V_B(s)$, in the column base is now seen as attributable to three factors: liquid flows entering and leaving the column base, and the two “swell factors”:

$$V_B(s) = \frac{w_{LR}(s) + w_1(s) - w_B(s) - w_v(s)}{\rho_L s} + k_{v1} w_{ST}(s) + k_{v2} H_B(s) \quad (16.36)$$

where

- w_1 = liquid from last downcomer due to reflux or feed, lbm/min
- w_{TR} = liquid downflow due to inverse or direct response, lbm/min
- w_B = bottom product flow, lbm/min
- w_v = vapor boilup, lbm/min

Inverse Response

As pointed out in Chapter 13, an *increase* in vapor flow up the column may cause the following:

1. A temporary *increase* in liquid flow down the column, which is called “inverse” response.
2. A temporary decrease in liquid flow down the column, which we term “direct” response.
3. No change in liquid downflow. This we call “neutral” response.

For either inverse or direct response, the mathematical relationship was shown to be:

$$\frac{w_{1R}(s)}{w_v(s)} \cong K_{TR} (1 - e^{-s\tau_{TR}}) \quad (16.37)$$

where K_{TR} has the units $\frac{\text{lbm/min}}{\text{lbm/min}} \times \text{min}$ and τ_{TR} is the individual tray hydraulic time constant in minutes. Note that when K_{TR}/τ_{TR} is both positive and greater than unity, there is inverse response for base level as well as for base composition. This is so because the reflux flow momentarily exceeds the boilup.

Overall Control System

If we are not limited in column-base holdup and can design for reasonably well-damped control, then we can treat reboiler dynamics as negligible. This says that steam flow responds to the flow controller set point immediately, and that boilup follows steam flow without lag. We may then prepare the signal flow diagram of Figure 16.7. Note that K_{mfs} = steam flow-meter gain = $12/(w_{ST})_{\max}$. This may be partially reduced to the form of Figure 16.8. From this last illustration we can see some of the loop’s characteristics as they are affected by reboiler swell and inverse response.

A. Reboiler Swell

From the denominator of the upper-right-hand term of Figure 16.8 we can see that for stability with proportional-only level control we must have:

$$A_B > \left(k_{v2} + k_{v1} \frac{K_{mh} K_{ch}}{K_{mfs}} \right) \quad (16.38)$$

The term k_{v2} is normally both negative and small. But as pointed out in Chapter 4, k_{v1} is often large at low heat loads; the control system designer must be careful about the term:

$$\frac{k_{v1} K_{mh} K_{ch}}{K_{mfs}}$$

At startup some control systems of this type will simply drive to a high base level and wide-open steam valve because of a large k_{v1} or improperly chosen values of the other parameters.

If a PI controller is used instead of a proportional-only controller, the denominator of the upper-right-hand box of Figure 16.8 becomes:

$$\tau_R (A_B - k_{v2})s - \frac{k_{v1} K_{mh} K_{ch} \tau_R s}{K_{mfs}} - \frac{k_{v1} K_{mh} K_{ch}}{K_{mfs}}$$

An examination of the mathematics suggests that for stability we should make $A_B - K_{v1} > k_{v1} K_{mh} K_{ch}/K_{mfs}$ and $\tau_R \lambda_\pi/\lambda_p > \rho_L k_{v1}$.

B. Inverse Response

If K_{TR} is positive, we have inverse response and the effect is unstablizing, as is swell.

A common approach to design, when holdup is adequately large, is to make loop calculations without the inverse response term and to design for a closed-loop natural period equal to or greater than $10n\tau_{TR}$.

If the control loop has to be tightly tuned because holdup is small, we can design a compensator for inverse response called an "inverse response predictor."⁸ It is analogous to the Smith predictor for dead-time compensation.⁹

The control loop containing the predictor is shown in Figure 16.9. The predictor contains a model of the inverse response and cancels its effect from the measured variable, allowing us to tune the controller as though inverse response were not present. The time-domain equation for the predictor is:

$$P(t) = K_{IP} [\theta_c(t - n \tau_{TR}) - \theta_c(t)] \Delta t + P(t - 1) \quad (16.39)$$

where

$$K_{IP} = \frac{\lambda_\pi}{\lambda_p} \times \frac{K_{mh}}{K_{mfs}} \frac{1}{\rho_L A_B} \times \frac{K_{TR}}{\tau_{TR}}$$

It can be shown by equation (16.39) that $P(t)$ will have a slight offset from set point. The impulse function shown in Figure 16.9 after the predictor eliminates this.

The inverse response predictor must be implemented in a microprocessor controller or control computer that has storage capability for the term $\theta_c(t - n \tau_{TR})$.

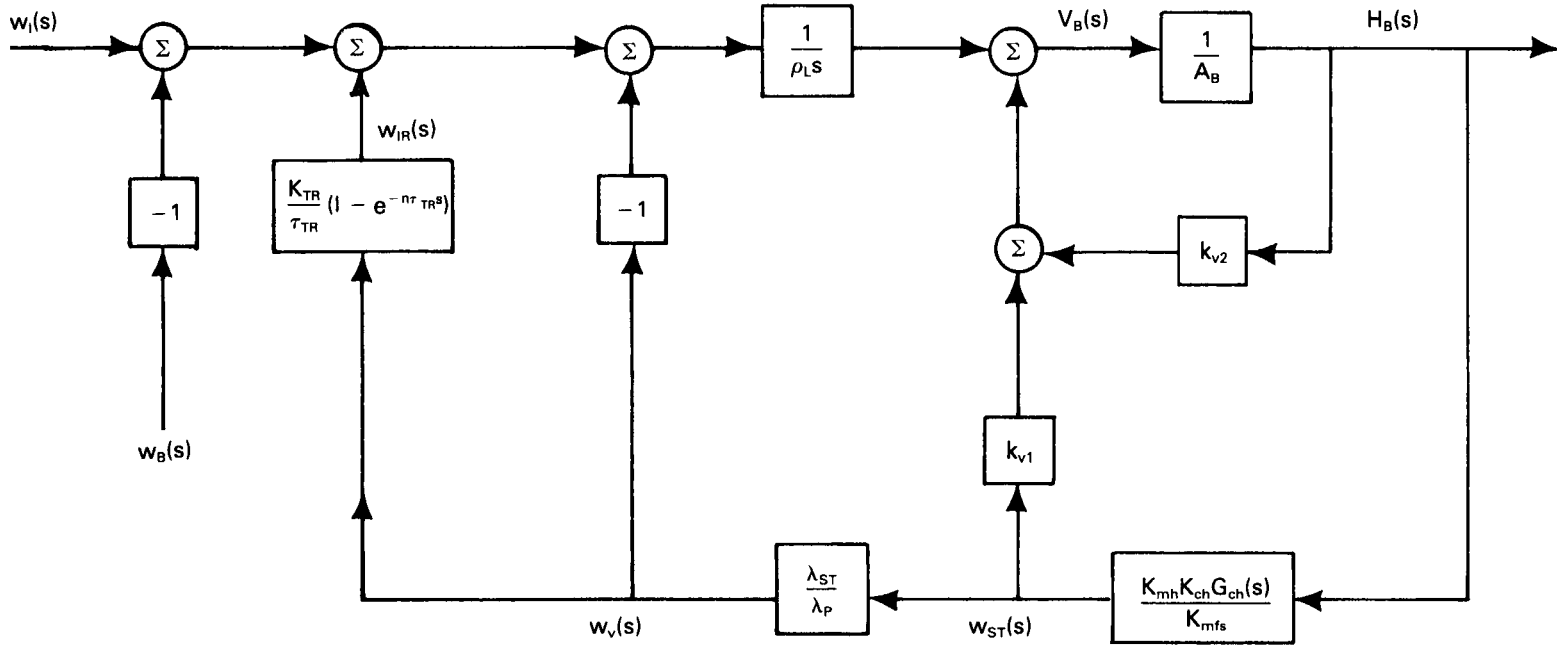


FIGURE 16.7
Base level control via steam flow control with inverse response and reboiler swell

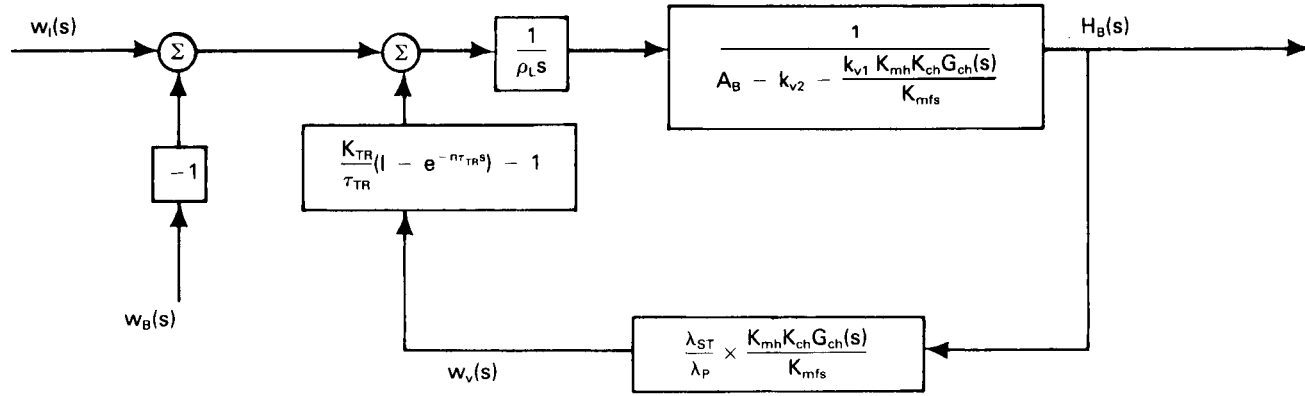


FIGURE 16.8
Partial reduction of figure 16.7

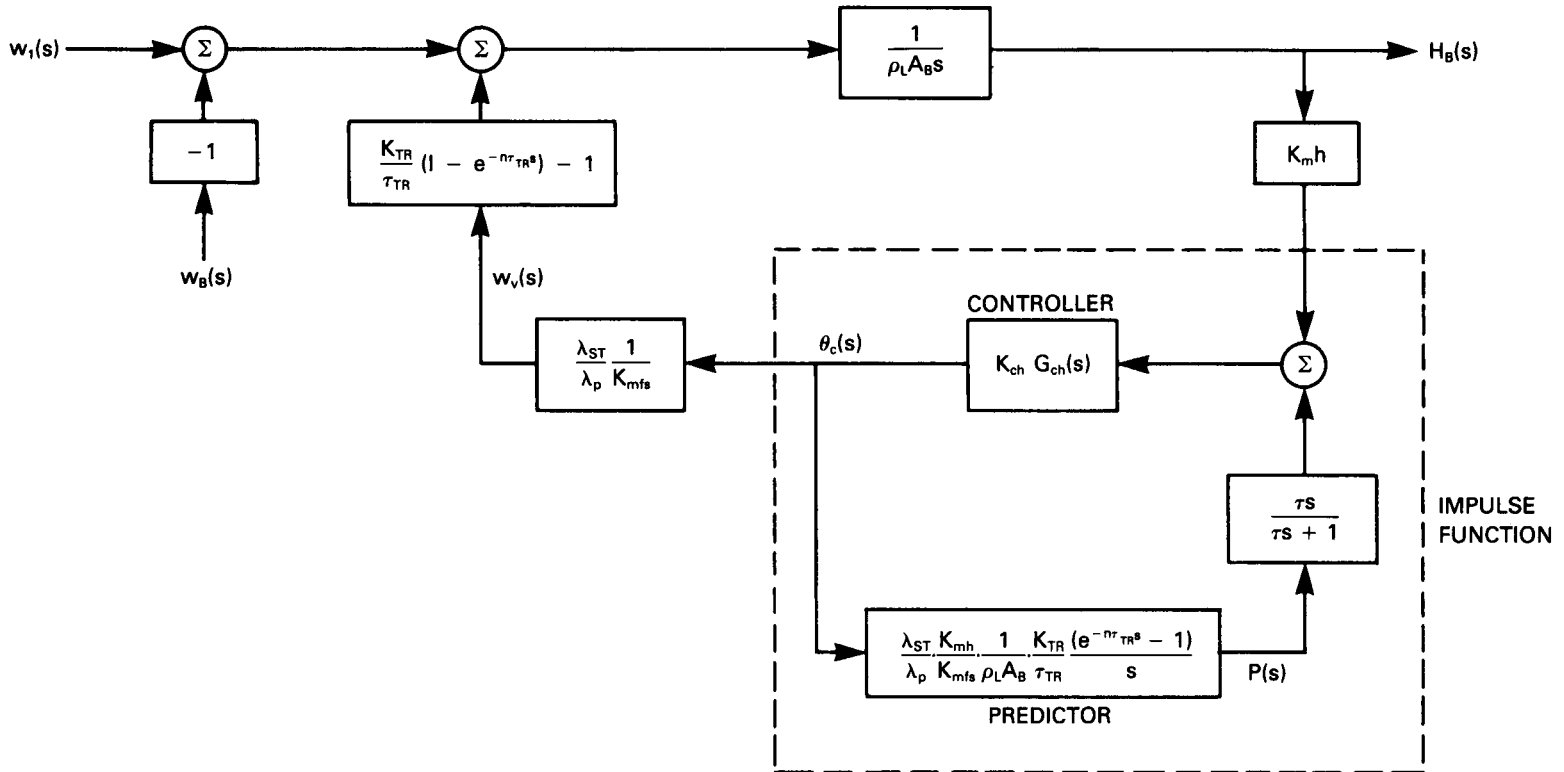


FIGURE 16.9
Inverse response predictor for base level control via steam flow

If K_{TR}/τ_{TR} is unity or close to it, the inverse response term in Figure 16.8 becomes simplified:

$$\frac{K_{TR}}{\tau_{TR}} (1 - e^{-\tau_{TR}s}) - 1 \cong -e^{-\tau_{TR}s} \quad (16.40)$$

The system now behaves as though it has dead time.

16.8 COLUMN-BASE LEVEL CONTROL VIA CONDENSATE THROTTLING FROM A FLOODED REBOILER (CASCADE LEVEL—FLOW CONTROL)

As a final example, let us look at the control of column-base level by throttling condensate flow from the reboiler. We will ignore possible column inverse response but will take swell into account. Column acoustic impedance, seen from the base, will be assumed to be resistive only. The basic, open-loop signal flow diagram for the flooded reboiler is given in Figure 15.13. We need in addition the following relationships:

$$w_o(s) = \frac{1}{K_{mf0}} \theta_{cb}(s) \quad (16.41)$$

$$w_v(s) = \frac{\lambda_{st}}{\lambda_p} w_\sigma(s) \quad (16.42)$$

$$V_B(s) = \frac{1}{\rho_L s} [w_1(s) - w_v(s) - w_B(s)] + k_{v1} w_v(s) \quad (16.43)$$

$$P_B(s) = \frac{R_{col}}{\rho_{BU}} w_v(s) \quad (16.44)$$

where

K_{mf0} = steam condensate flow-meter gain

$$= 12/(w_o)_{\max}$$

w_o = steam condensate flow, lbm/min

w_v = boilup, lbm/min

w_σ = rate of steam condensation, lbm/min

V_B = liquid volume between level taps, ft³

ρ_L = liquid density, lbm/ft³

w_1 = liquid downflow from column, lbm/min

k_{v1} = swell coefficient, $\frac{\text{ft}^3 \text{ liquid}}{\text{lbm/min boilup}}$

R_{col} = column acoustic resistance, lbf min/ft⁵.

It is assumed that column impedance is fast enough to reduce to a constant.

P_B = column-base pressure, lbf/ft²

ρ_{BU} = process vapor density, lbm/ft³

For a small number of flooded reboilers examined to date, the k_{v1} term is essentially constant from low to high heat loads; that is, the relationship between heat load and tube vapor volume is a straight line. The compression effect due to process-side liquid-elevation changes has not been checked but is believed to be small; it is neglected in this analysis.

The preliminary signal flow diagram of Figure 16.10 may now be prepared. It may be reduced readily to the form of Figure 16.11.

Some consolidation of terms of Figure 16.11 may now be accomplished:

$$k_{v1} - \frac{1}{\rho_L s} = \frac{k_{v1} \rho_L s - 1}{\rho_L s} \quad (16.45)$$

$$= \frac{K_1 (\tau_1 s - 1)}{s} \quad (16.46)$$

Next, let:

$$\mu(s) = \frac{\Omega(s) \lambda_{st}/\lambda_p}{1 + \Omega(s) \frac{\lambda_{st}}{\lambda_p} \bar{U}_R \bar{A}_c \frac{\partial T_{BU}}{\partial P_B} \frac{R_{col}}{\rho_{BU}}} \quad (16.47)$$

From equation (15.27a):

$$\Omega(s) = \frac{\bar{V}_{st}/\bar{P}_\alpha \lambda_{st}}{\alpha \bar{V}_{st}/\bar{P}_\alpha + \frac{\bar{U}_R \bar{A}_c}{\rho_{st} \lambda_{st}} \frac{\partial T_\alpha}{\partial P_\alpha}} \times \frac{s}{\frac{\bar{V}_{st}/\bar{P}_\alpha}{\alpha \bar{V}_{st}/\bar{P}_\alpha + \frac{\bar{U}_R \bar{A}_c}{\rho_{st} \lambda_{st}} \frac{\partial T_\alpha}{\partial P_\alpha}} s + 1} \quad (16.48)$$

$$= \frac{K''_{FC} s}{\tau'_{FC} s + 1} \quad (16.49)$$

so

$$\mu(s) = \frac{\frac{K''_{FC} s}{\tau'_{FC} s + 1} \times \frac{\lambda_{st}}{\lambda_p}}{1 + \frac{K''_{FC} s}{(\tau'_{FC} s + 1)} \left(\frac{\lambda_{st} \bar{U}_R \bar{A}_c}{\lambda_p} \right) \frac{\partial T_{BU}}{\partial P_B} \frac{R_{col}}{\rho_{BU}}} \quad (16.50)$$

$$= \frac{(\lambda_{st}/\lambda_p) K''_{FC} s}{\left(\tau'_{FC} + K''_{FC} \frac{\lambda_{st} \bar{U}_R \bar{A}_c}{\lambda_p} \times \frac{\partial T_{BU}}{\partial P_B} \times \frac{R_{col}}{\rho_{BU}} \right) s + 1} \quad (16.51)$$

$$= \frac{(\lambda_{st}/\lambda_p) K''_{FC} s}{\tau_2 s + 1} \quad (16.52)$$

From Figure 16.11 and equations (16.48) and (16.52) we can see that the characteristic equation for this system is:

$$1 - \frac{(\lambda_s/\lambda_p) K_{FC}'' s}{\tau_2 s + 1} \times \frac{K_1 (\tau_1 s - 1)}{s} \times \frac{K_{mh} K_{ch} G_{ch}(s) \alpha \lambda_s}{A_B K_{mfo} s} = 0 \quad (16.53)$$

For proportional-only control this reduces to:

$$1 - \frac{K_L (\tau_1 s - 1)}{s (\tau_2 s + 1)} = 0 \quad (16.54)$$

or

$$\tau_2 s^2 + (1 - K_L \tau_1) s + K_L = 0 \quad (16.55)$$

or

$$\frac{\tau_2}{K_L} s^2 + \left(\frac{1 - K_L \tau_1}{K_L} \right) s + 1 = 0 \quad (16.56)$$

It can be seen from equation (16.56) that for stability we must make $K_L \tau_1 \leq 1$. But this is not enough for a practical design; the damping ratio should be at least one, where:

$$\frac{1 - K_L \tau_1}{K_L} = 2 \zeta \sqrt{\frac{\tau_2}{K_L}} \quad (16.57)$$

or

$$\zeta = \frac{1 - K_L \tau_1}{2 K_L} \sqrt{\frac{K_L}{\tau_2}} = \frac{1 - K_L \tau_1}{2 \sqrt{K_L \tau_2}} \quad (16.58)$$

For $\zeta = 1$:

$$4 K_L \tau_2 = 1 - 2 K_L \tau_1 + K_L^2 \tau_1^2 \quad (16.59)$$

Now let

$$\tau^2 = k_R \tau_1$$

Then:

$$4 K_L k_R \tau_1 = 1 - 2 K_L \tau_1 + K_L^2 \tau_1^2 \quad (16.60)$$

or

$$1 - 2 K_L \tau_1 (2 k_R + 1) + K_L^2 \tau_1^2 = 0 \quad (16.61)$$

whence

$$K_L = \frac{(2 k_R + 1) \pm 2 \sqrt{k_R^2 + k_R}}{\tau_1} \quad (16.62)$$

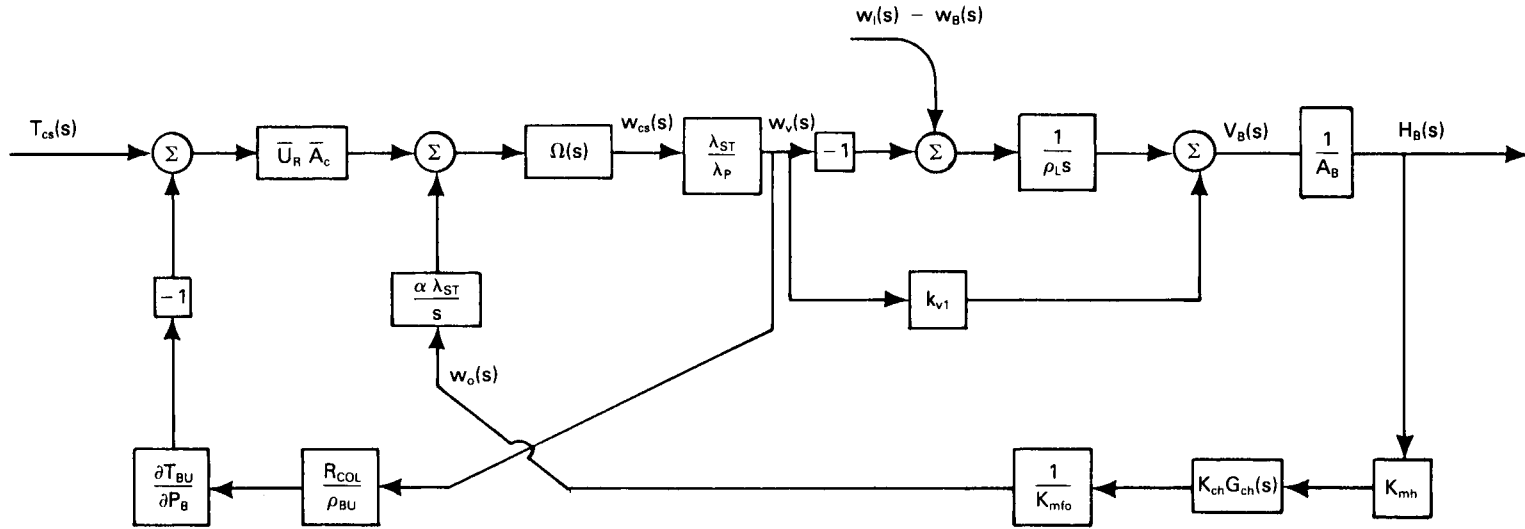


FIGURE 16.10
Preliminary signal flow diagram for column base level control via condensate throttling from a flooded reboiler

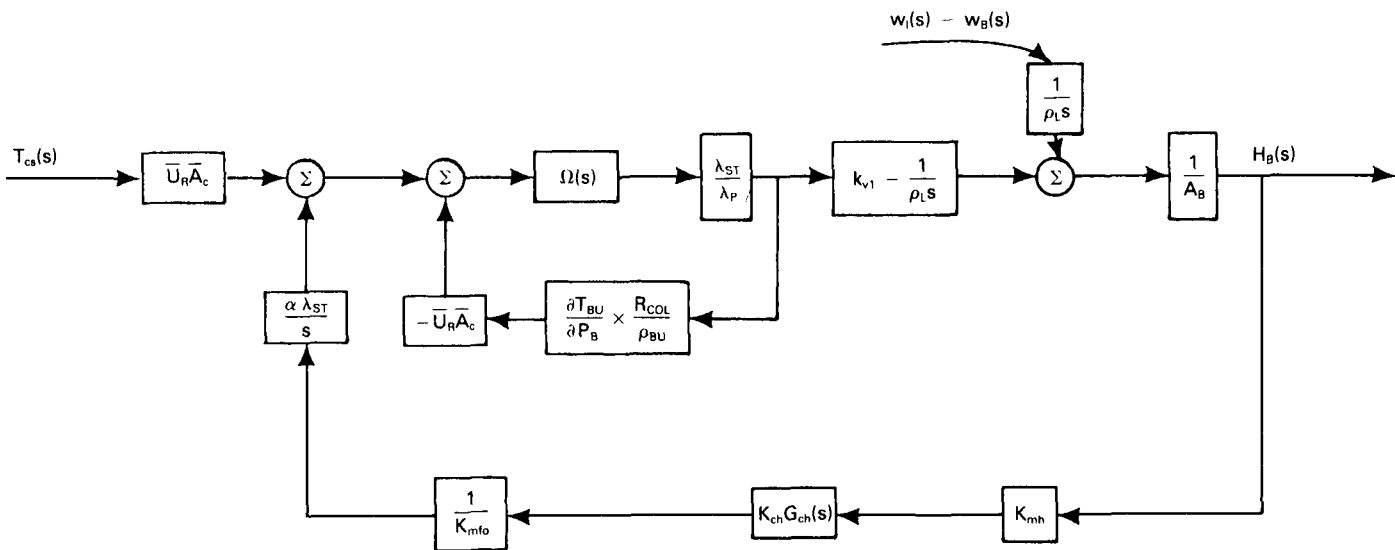


FIGURE 16.11
Reduced version of figure 16.10

Since for slow, well-damped response we want $\tau_Q = \tau_2/K_L$ to be as large as possible, we should choose from equation (16.62) the solution that gives the smallest value of K_L . For a proportional-reset controller, choose:

$$\tau_R \cong 5 \tau_Q = 5 \sqrt{\frac{\tau_2}{K_L}} \quad (16.63)$$

With such a controller it will be necessary, as indicated earlier, to use auto overrides or a controller with nonlinear gain and reset.

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17 Pressure and ΔP Control

17.1 INTRODUCTION

Column pressure control, as mentioned earlier, usually can be of the “averaging” type. Most columns do not need “tight” pressure control, and for some of them it is undesirable; a rapid change in pressure can cause flashing or cessation of boiling in the column. The former might result in flooding and the latter in dumping. For well-damped pressure control, the mathematical models in this chapter, which mostly treat condenser and reboiler dynamics as negligible, are usually adequate.

For tight pressure control, we should use these models with caution. Most of the tight column pressure controls we have studied have closed-loop resonant frequencies in the range of 0.5–2 cpm. For the upper value one should make at least a rough check of condenser and reboiler dynamics. It may be of interest that the only applications of tight pressure control we have found are in heat-recovery schemes where the vapor from one column serves as the heating medium for the reboiler of another column, and perhaps furnishes heat to other loads. If the vapor flow must be throttled to each load, constant up- and downstream pressures help good flow control.

17.2 HEAT-STORAGE EFFECT ON COLUMN PRESSURE

The stored heat in the liquid in the column and its base can exercise quite a leveling effect on pressure and differential pressure. If pressure starts to drop, more liquid flashes, which tends to reduce the rate at which pressure falls. If pressure starts to rise, the reverse happens—the rate of boiling is decreased and pressure rises at a slower rate. In this section we will see how vaporization rate, w_v , is affected by heat storage.

A simplified analysis can be made if:

1. Column ΔP is small compared with absolute pressure.
2. Latent heats at the top and bottom of the column are nearly the same.
3. And the temperature difference across the column is not too large.

These assumptions permit us, at least for atmospheric or pressurized columns, to lump the column contents together and to assume an average temperature, T_{CP} , and an average pressure, P_{CP} , as shown by Figure 17.1. Note that the reboiler is lumped with the column base but the condenser and overhead receiver are not lumped with the column top.

We may now write a heat-balance equation similar to equation (15.30) but with terms for reflux added:

$$\begin{aligned} c_p \bar{T}_F \bar{w}_F(s) + c_p \bar{w}_F T_F(s) + q_T(s) - c_p [(\bar{w}_c + \bar{w}_B) + s W_{col}] T_{cp}(s) \\ + c_p \bar{T}_R \bar{w}_R(s) + c_p \bar{w}_R T_R(s) - c_p \bar{T}_{cp} \bar{w}_B(s) - c_p s \bar{T}_{cp} W_{col}(s) \quad (17.1) \\ = (\lambda_p + c_p \bar{T}_{cp}) \bar{w}_c(s) \end{aligned}$$

The material-balance equation is:

$$W_{col}(s) = \frac{1}{s} [w_F(s) + w_R(s) - w_c(s) - w_B(s)] \quad (17.2)$$

Reboiler with Steam Flow or Flow-Ratio Controlled

If we treat reboiler dynamics as negligible, and if we assume that steam will be flow or flow-ratio controlled, we may combine the above equations as shown in the preliminary signal flow diagram of Figure 17.2. This, in turn, may be reduced to the form of Figure 17.3.

Reboiler with Direct Throttled Steam

For the case where the steam valve is manipulated by some variable other than flow or flow ratio, we may need to account for reboiler dynamics to calculate q_T . Referring to Figure 15.8, we may make a partial signal flow diagram as shown on Figure 17.4 where:

- C_R = reboiler hot-side acoustic capacitance, ft^5/lbf
- A_R = reboiler heat-transfer area, ft^2
- P_α = reboiler hot-side pressure, lbf/ft^2
- Q_r = heating-medium flow, actual ft^3/sec
- ρ_{sr} = heating-medium (vapor) density, lbm/ft^3
- λ_{sr} = heating-medium latent heat, pcu/lb
- U_R = reboiler heat-transfer coefficient, $\frac{\text{pcu}/\text{sec}}{^\circ\text{C ft}^2}$

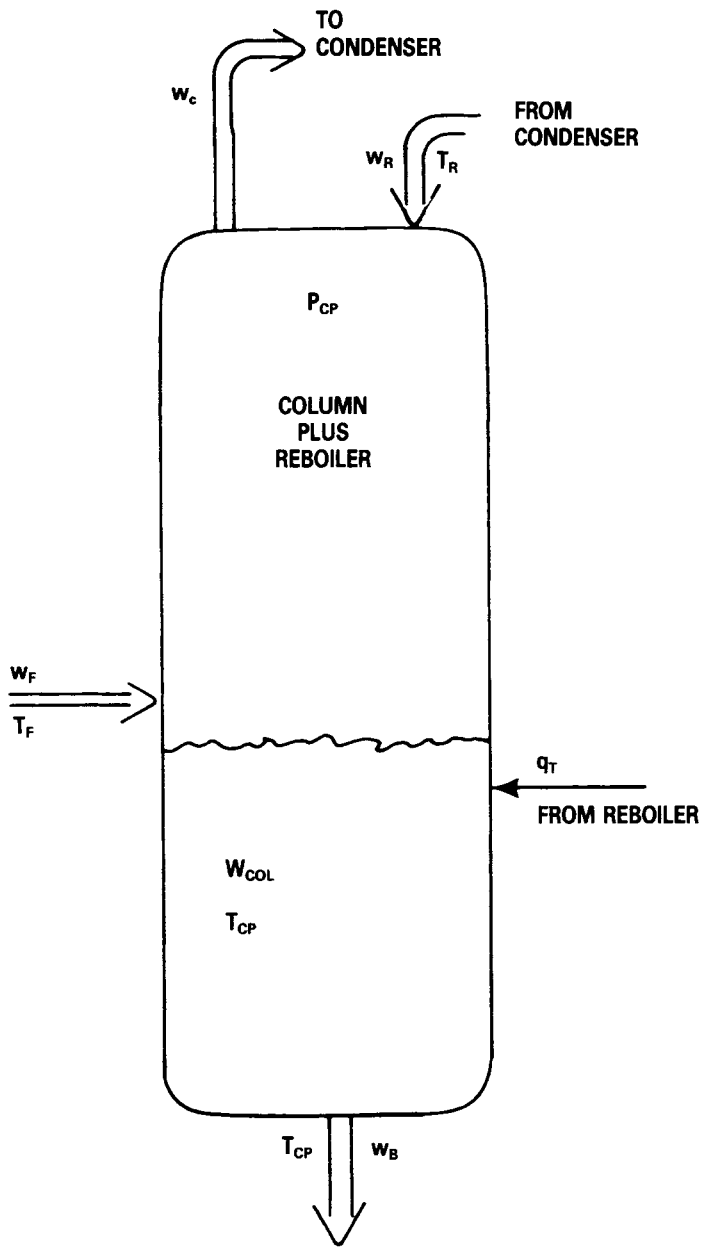


FIGURE 17.1
Simplified treatment of heat storage effect on column pressure dynamics

This may be reduced to the form of Figure 17.5 where:

$$\beta(s) = \frac{\bar{U}_R A_r}{1 + \frac{\bar{U}_R A_R}{\lambda_{st} \rho_{st}} \times \frac{\partial T_{\alpha} / \partial P_{\alpha}}{C_{R^s} - \partial Q_v / \partial P_{\alpha}}} \quad (17.3)$$

For critical flow, note that $\partial Q_v / \partial P_{\alpha} = 0$.

Figures 17.3 and 17.5 may now be combined into the signal flow diagram of Figure 17.6.

17.3 PRESSURE CONTROL VIA VENT AND INERT GAS VALVES

As mentioned in Chapter 3, we often control pressure in a column by a pressure-dividing network such as shown in Figure 3.7. The control-valve input-signal spans are usually the same but one valve opens while the other closes.

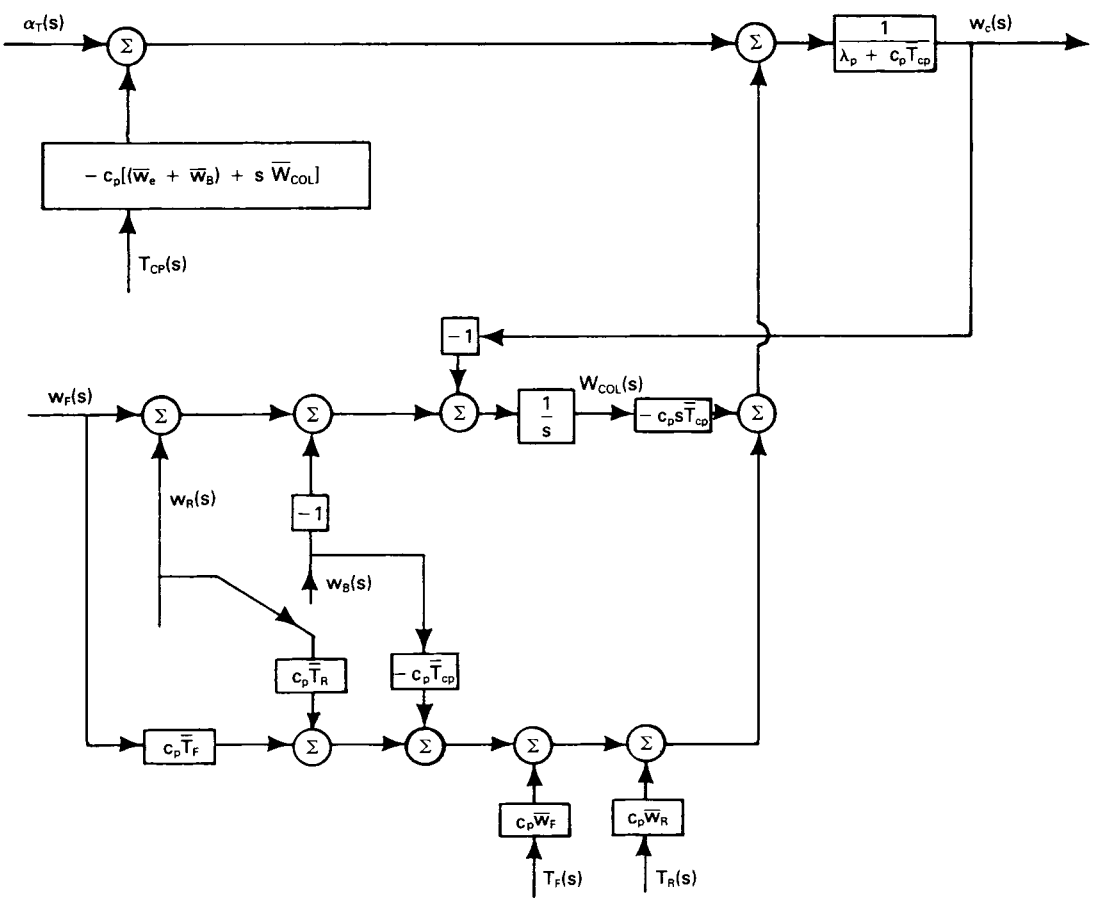


FIGURE 17.2
Preliminary signal flow diagram for column heat storage dynamics

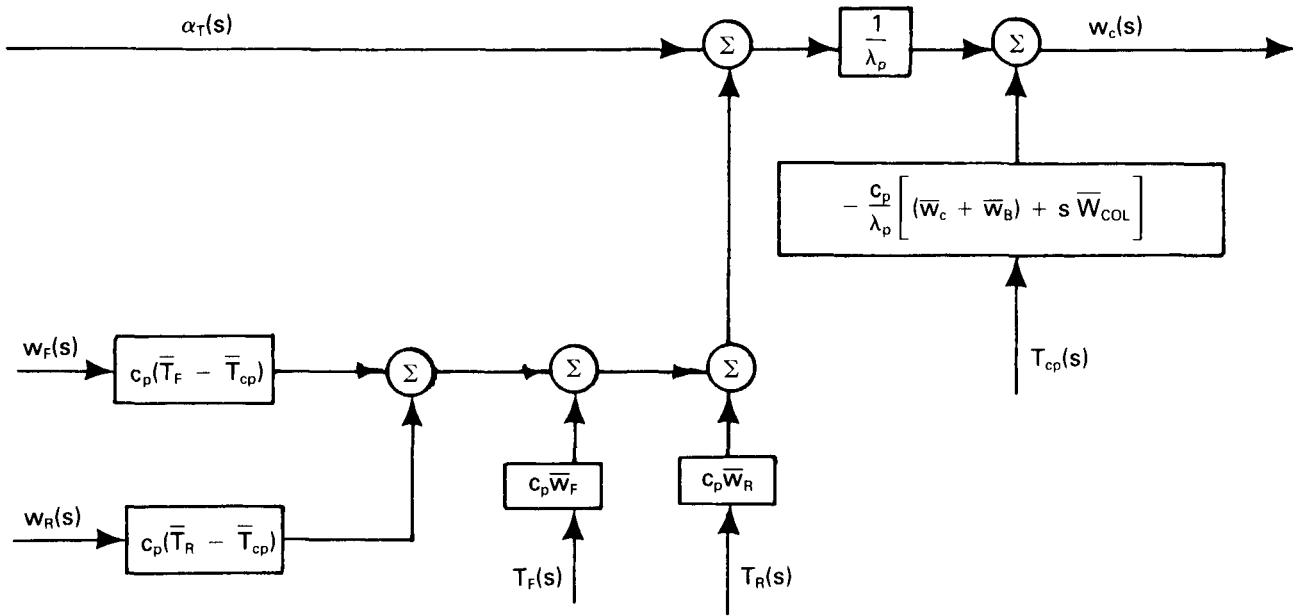


FIGURE 17.3
Reduced signal flow diagram of preliminary diagram

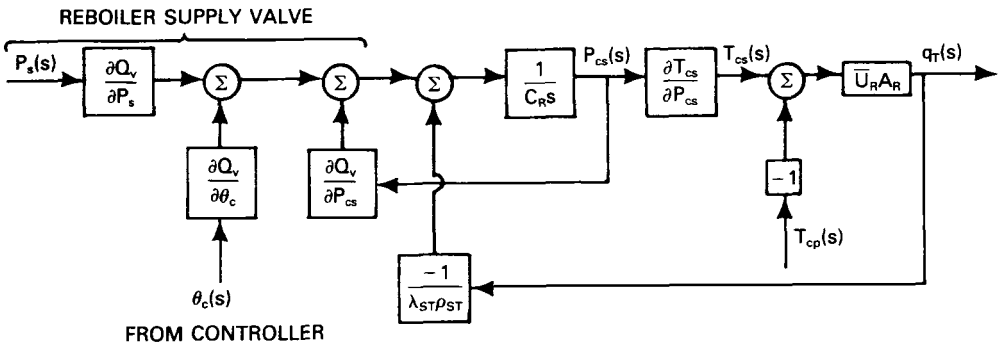


FIGURE 17.4
Partial signal flow diagram for reboiler dynamics

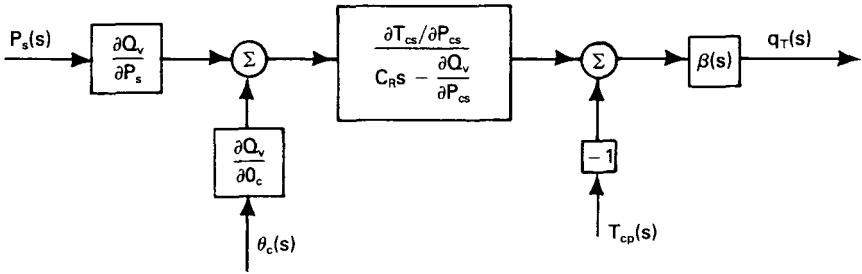


FIGURE 17.5
Reduced form of signal diagram of figure 17.4

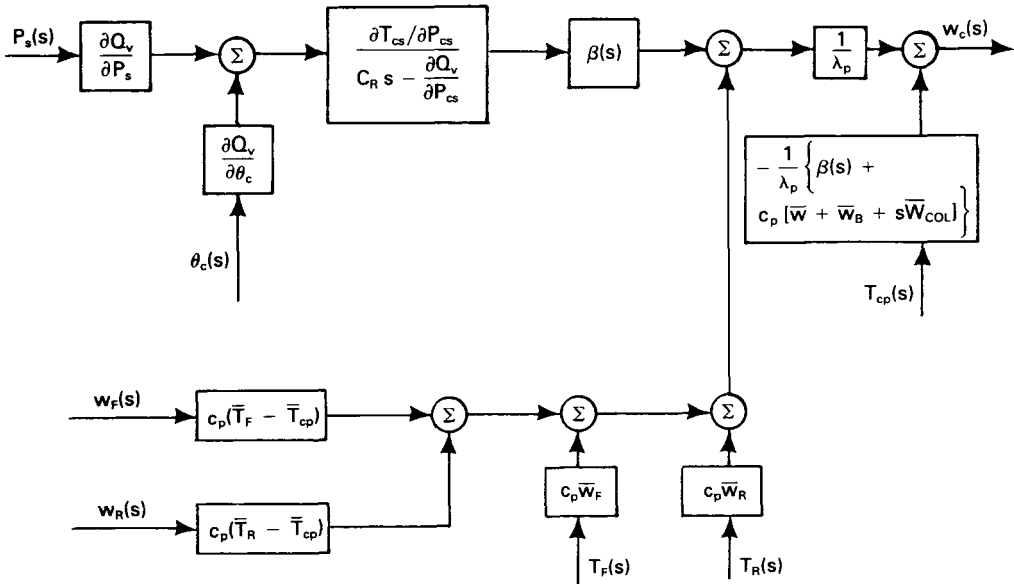


FIGURE 17.6
Combined signal flow diagram for figures 17.3 and 17.5

The equation for the inert gas valve is:

$$w_{IG}(s) = \frac{\partial w_{IG}}{\partial P_{IG}} P_{IG}(s) + \frac{\partial w_{IG}}{\partial P_{\varphi}} P_{\varphi}(s) + \frac{\partial w_{IG}}{\partial \theta_c} \theta_c(s) \quad (17.4)$$

Similarly, the equation for the vent valve is:

$$w_c(s) = \frac{\partial w_c}{\partial P_{\varphi}} P_{\varphi}(s) + \frac{\partial w_c}{\partial P_R} P_R(s) + \frac{\partial w_c}{\partial \theta_c} \theta_c(s) \quad (17.5)$$

where

w_{IG} = inert gas flow, lbm/sec

w_c = vent flow, lbm/sec

P_{φ} = column pressure, lbf/ft²

P_{IG} = inert gas supply pressure, lbf/ft²

P_R = pressure downstream of vent valve, lbf/ft²

θ_c = controller output signal

These equations may be combined into the partial signal flow diagram of Figure 17.7. Note that the valve gains, $\partial w_{IG}/\partial \theta_c$ and $\partial w_c/\partial \theta_c$, are both assumed to be positive; reverse action of the inert gas valve is obtained by the -1 term. If we can assume that P_{IG} and P_R are sufficiently constant, we may reduce Figure 17.7 to the form of Figure 17.8. The term C_c is the acoustic capacitance of the column, vapor line to condenser, and the condenser.

For the case where reboiler steam is flow or flow-ratio controlled we can now combine Figures 17.8 and 17.3 into the signal flow diagram of Figure 17.9. Note the pressure feedback on w_c through $\partial T_{\varphi}/\partial P_{\varphi}$, and the addition of the pressure measurement, $K_m G_m(s)$, and the controller, $K_c G_c(s)$. Figure 17.9 can be reduced to the form of Figure 17.10 where:

$$\begin{aligned} K_{P1} G_{P1}(s) = & \frac{1}{\left(\frac{\partial w_c}{\partial P_{\varphi}} - \frac{\partial w_{IG}}{\partial P_{\varphi}} \right) + \frac{\partial T_{\varphi}}{\partial P_{\varphi}} \frac{c_p}{\lambda_p} (\bar{w}_c + \bar{w}_B)} \\ & \times \frac{1}{\rho_v C_c + \frac{\partial T_{\varphi}}{\partial P_{\varphi}} \frac{c_p}{\lambda} W_{col}} s + 1 \\ & \left(\frac{\partial w_c}{\partial P_{\varphi}} - \frac{\partial w_{IG}}{\partial P_{\varphi}} \right) + \frac{\partial T_{\varphi}}{\partial P_{\varphi}} \frac{c_p}{\lambda_p} (\bar{w}_c + \bar{w}_B) \end{aligned} \quad (17.6)$$

From equation (17.6) we can see that open-loop pressure dynamics are essentially first order. Since in most cases the inert gas bleed and the vent flow are fairly small, the valve gains, $\partial w_{IG}/\partial \theta_c$ and $\partial w_c/\partial \theta_c$, tend to be small. Together with the first-order dynamics, this commonly leads to large controller gains (small proportional bands) and control valve saturation for fairly small disturbances.

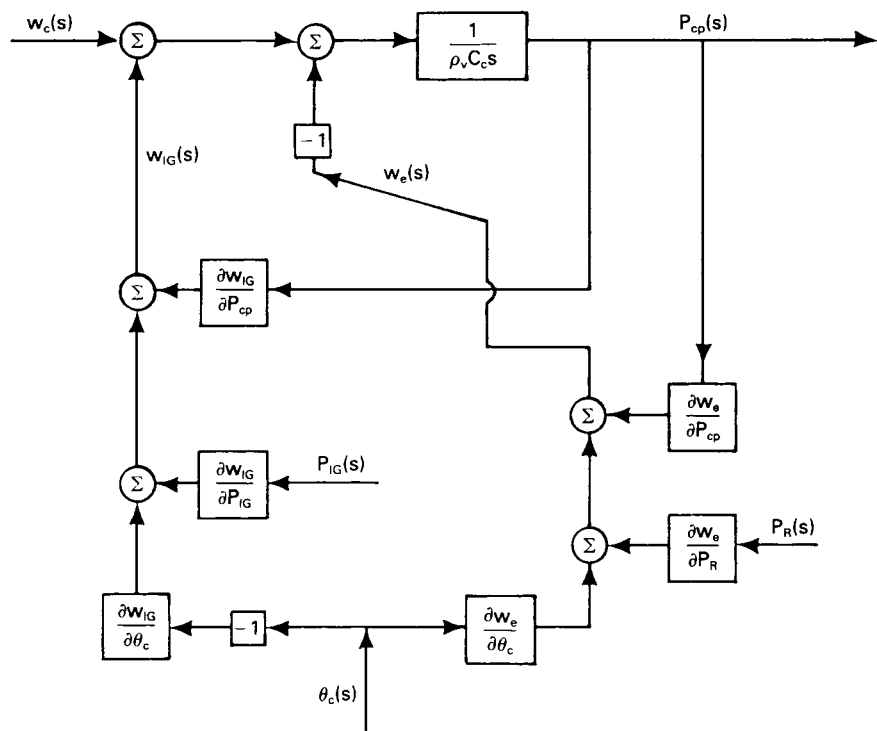


FIGURE 17.7
Partial signal flow diagram for column pressure control via manipulation of iner gas and vent valves

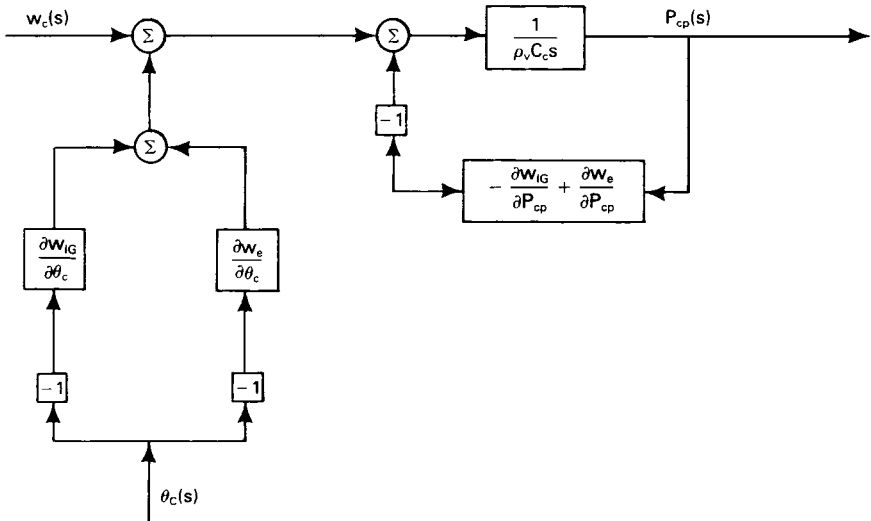


FIGURE 17.8
Reduction of signal flow diagram of figure 17.7

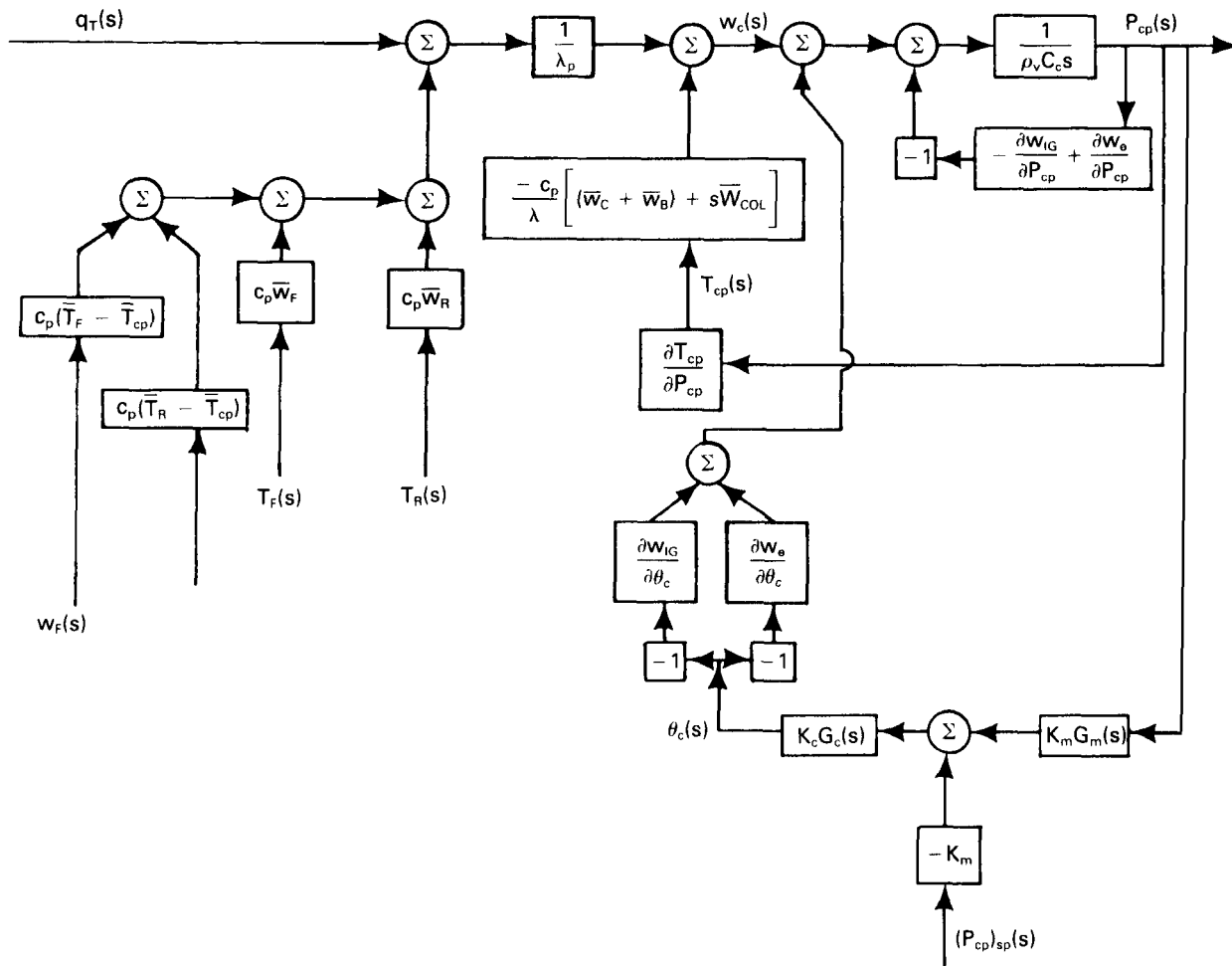


FIGURE 17.9

Signal flow diagram for column pressure control via manipulation of inert gas and vent valves when reboiler steam is flow or flow ratio controlled

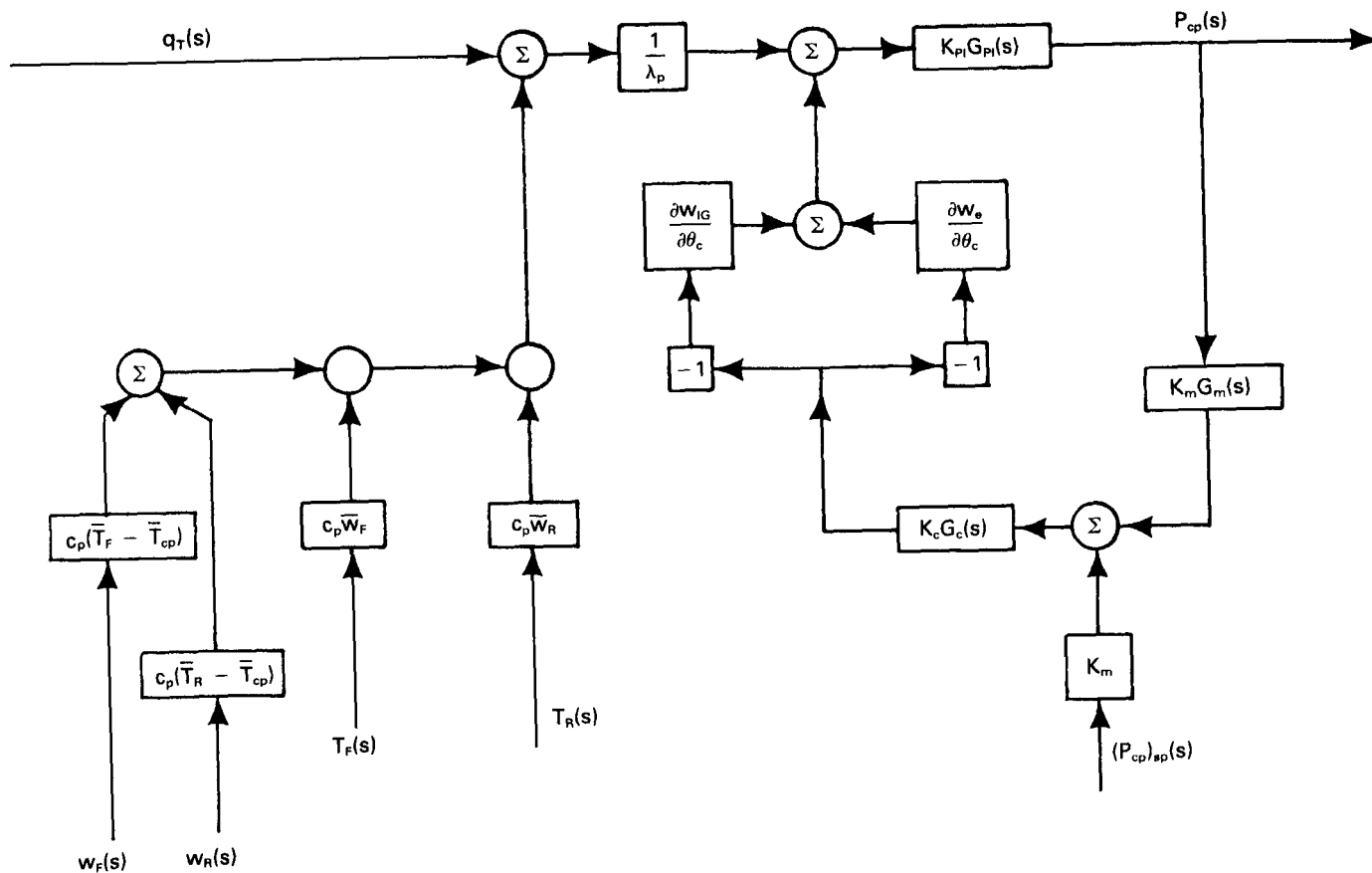


FIGURE 17.10
Reduced form of figure 17.9

17.4 PRESSURE CONTROL VIA FLOODED CONDENSER

Here we consider four cases:

1. Flow- or flow-ratio-controlled steam to reboiler, significant inerts.
2. Flow- or flow-ratio-controlled steam to reboiler, negligible inerts.
3. Steam to reboiler not flow or flow-ratio controlled, significant inerts.
4. Steam to reboiler not flow or flow-ratio controlled, negligible inerts.

Flow- or Flow-Ratio-Controlled Reboiler Steam, Significant Inerts

For this case we combine Figures 15.4 and 17.3, which leads to the signal flow diagram of Figure 17.11. This reduces to the form of Figure 17.12 where:

$$K_{P2}G_{P2}(s) = \frac{\frac{K_{Fc}(s + \alpha)}{\tau_Q^2 s^2 + 2 \zeta \tau_Q s + 1}}{1 + \frac{K_{Fc}(s + \alpha)}{\tau_Q^2 s^2 + 2 \zeta \tau_Q s + 1} \times \frac{c_p}{\lambda_p} [(\bar{w}_e + \bar{w}_B) + s \bar{W}_{col}] \frac{\partial T_{\varphi}}{\partial P_{\varphi}}} \quad (17.7)$$

This will clear to an expression that is first order in the numerator and second order in the denominator.

Flow- or Flow-Ratio-Controlled Steam, Negligible Inerts

This system may also be represented by a signal flow diagram similar to that of Figure 17.12 except that $K_{P3}G_{P3}(s)$ replaces $K_{P2}G_{P2}(s)$ and has a slightly different definition [see equation (15.26)]:

$$K_{P3}G_{P3}(s) = \frac{\frac{K'_{Fc}(s + \alpha)}{s(\tau'_{Fc} s + 1)}}{1 + \frac{K'_{Fc}(s + \alpha)}{s(\tau'_{Fc} s + 1)} \times \frac{c_p}{\lambda_p} [(\bar{w}_e + \bar{w}_B) + s \bar{W}_{col}] \frac{\partial T_{\varphi}}{\partial P_{\varphi}}} \quad (17.8)$$

Steam to Reboiler Not Flow or Flow-Ratio Controlled, Significant Inerts

Pressure controls via a flooded condenser with significant inerts and where reboiler steam is not flow or flow-ratio controlled may be represented by a signal flow diagram (Figure 17.13) formed by the combination of Figures 17.6 and 15.4. This diagram may be reduced to the form of Figure 17.14 where:

$$\begin{aligned} & K_{P4}G_{P4}(s) \\ &= \frac{\frac{K_{Fc}(s + \alpha)}{\tau_Q^2 s^2 + 2 \zeta \tau_Q s + 1}}{1 + \frac{K_{Fc}(s + \alpha)}{\tau_Q^2 s^2 + 2 \zeta \tau_Q s + 1} \times \frac{\partial T_{\varphi}}{\partial P_{\varphi}} \times \frac{1}{\lambda_p} [\beta(s) + c_p (\bar{w}_e + \bar{w}_B) + s \bar{W}_{col}]} \end{aligned} \quad (17.9)$$

Note that the numerator comes from equation (15.25) while $\beta(s)$ is given by equation (17.3).

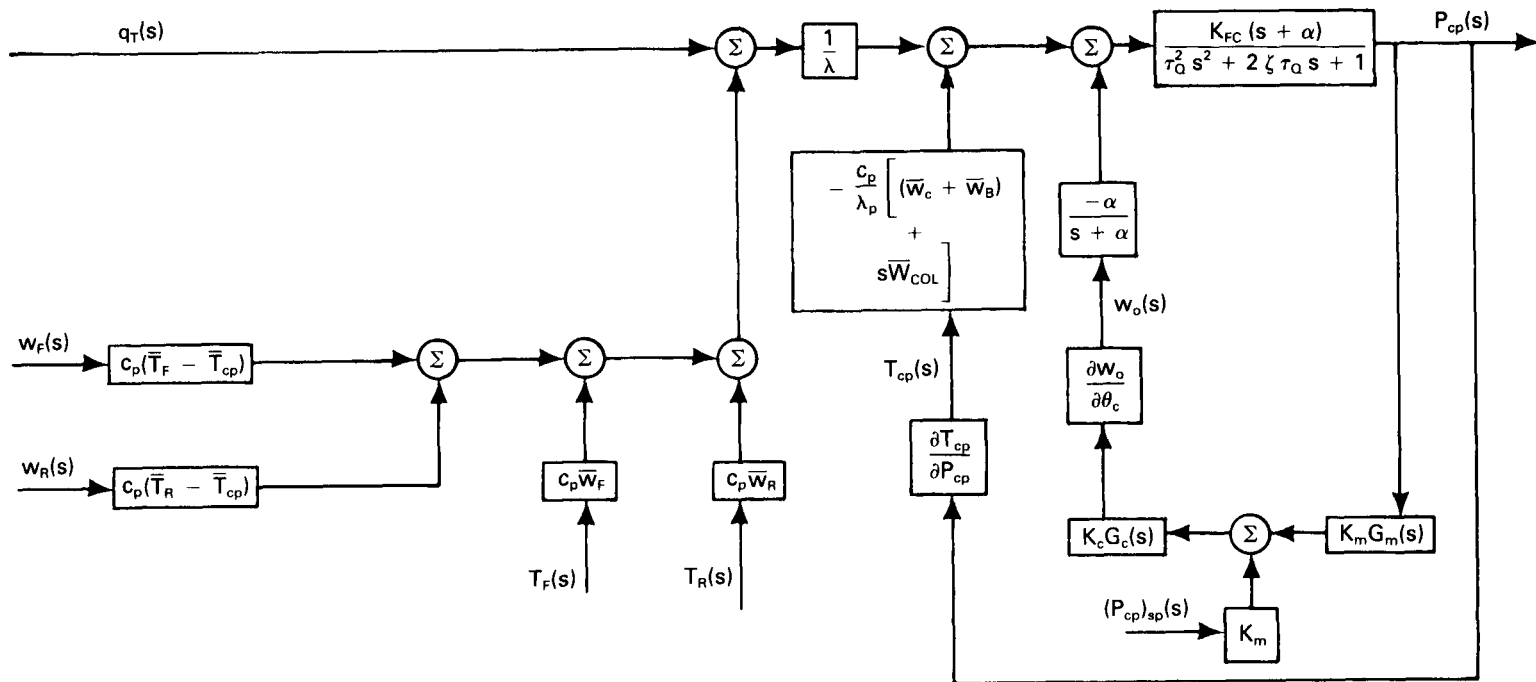


FIGURE 17.11

Column pressure control via flooded condenser drain—negligible inerts and reboiler steam flow or flow—ratio controlled

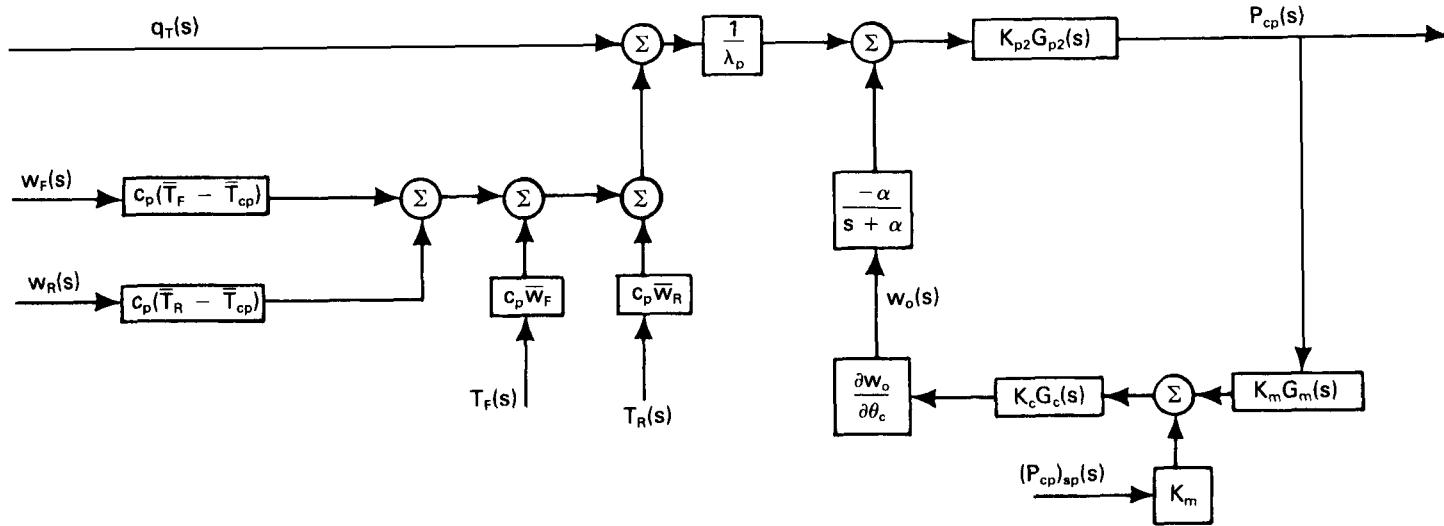


FIGURE 17.12
Partially reduced version of figure 17.11

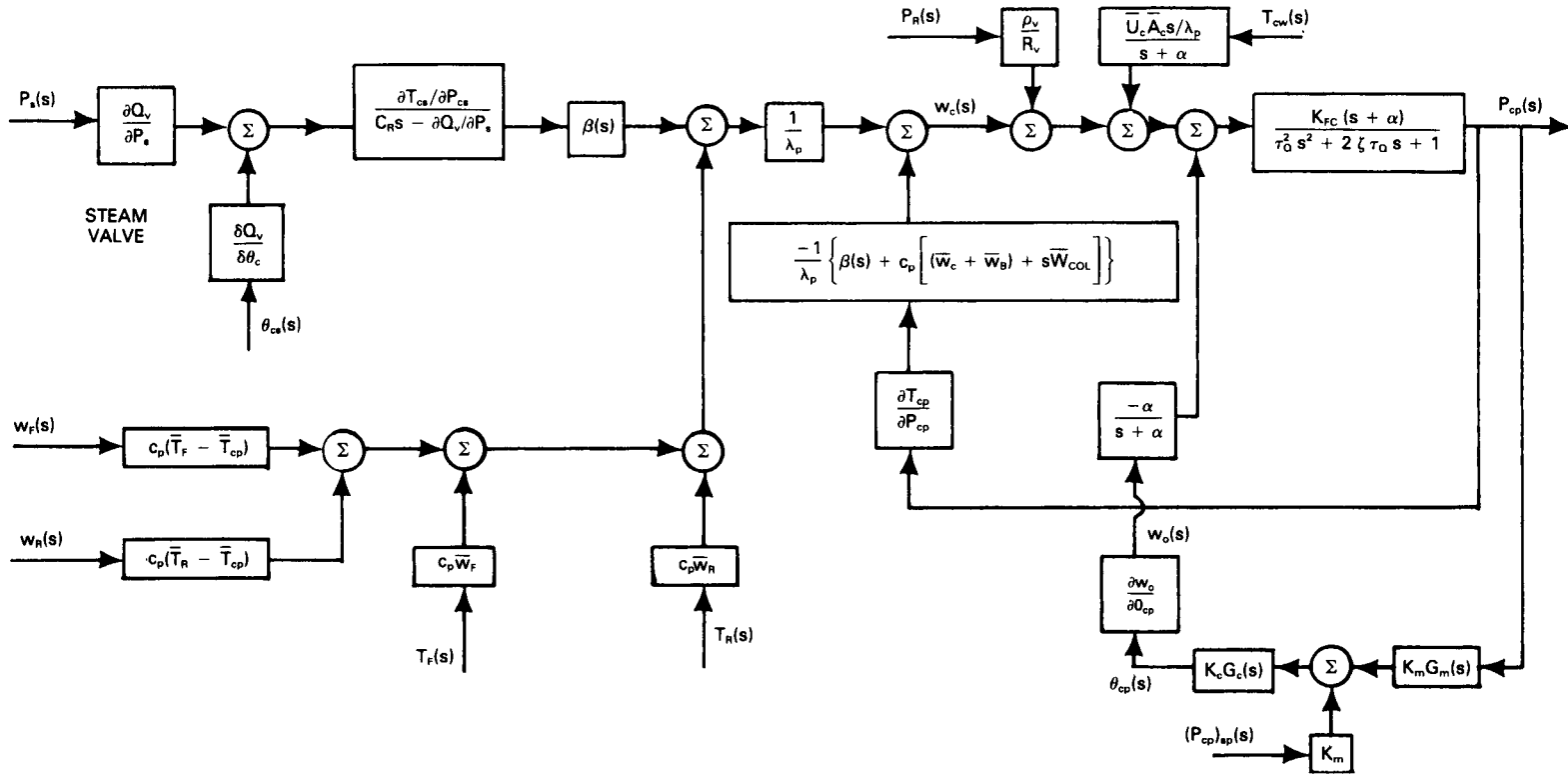


FIGURE 17.13
 Column pressure control via flooded condenser, reboiler steam
 not flow or flow ratio controlled, significant inerts

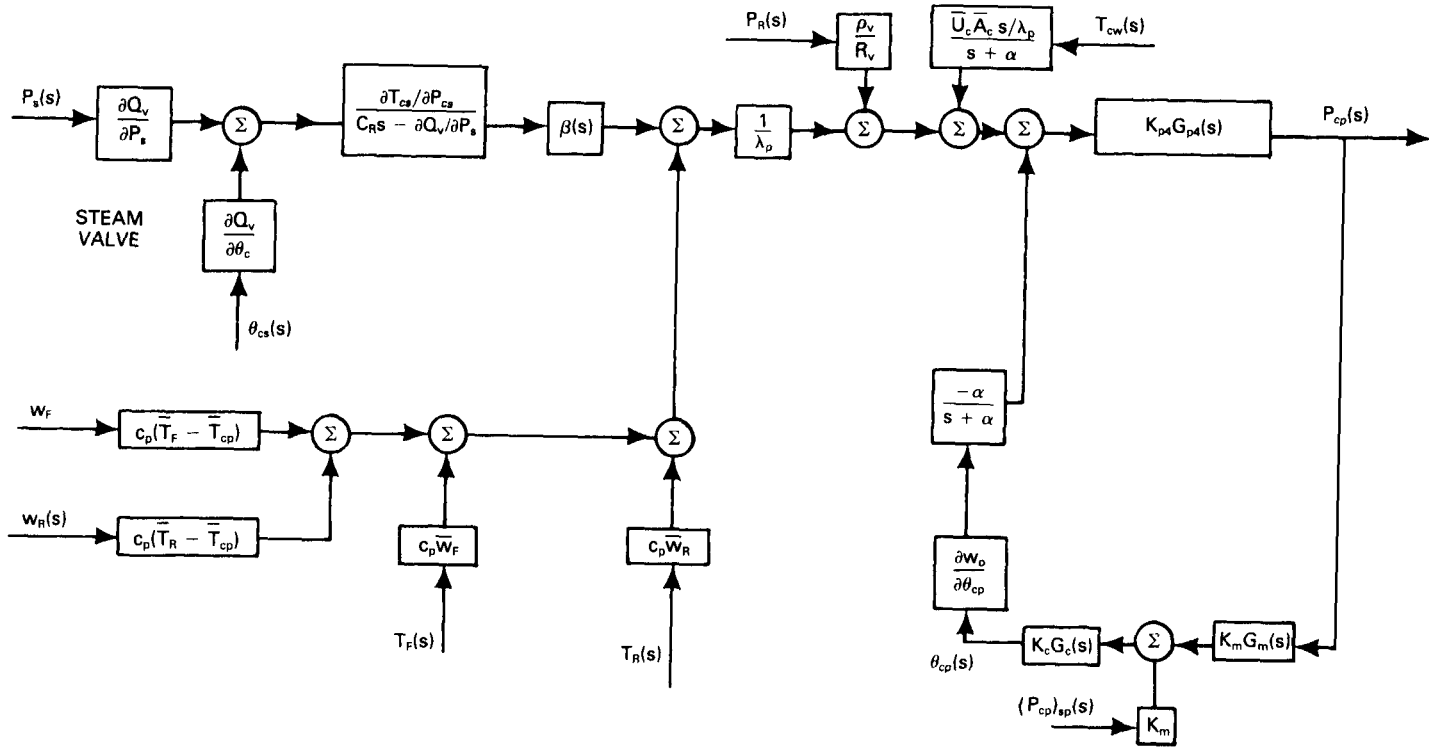


FIGURE 17.14
Partially reduced version of figure 17.13

Steam to Reboiler Not Flow or Flow-Ratio Controlled, Negligible Inerts

The signal flow diagram for this case is the same as Figure 17.14 except that $K_{P5}G_{P5}(s)$ is defined differently [see equation (13.26)]:

$$K_{P5}G_{P5}(s) = \frac{\frac{K'_{Fc}(s + \alpha)}{s(\tau'_{Fc}s + 1)}}{1 + \frac{K'_{Fc}(s + \alpha)}{s(\tau'_{Fc}s + 1)} \times \frac{\partial T_{\varphi}}{\partial P_{\varphi}} \times \frac{1}{\lambda_p} (\beta(s) + c_p[(\bar{w}_e + \bar{w}_B) + s\bar{W}_{col}]})} \quad (17.10)$$

17.5 PRESSURE CONTROL VIA CONDENSER COOLING WATER

This method of controlling pressure, although once popular, has fallen into some disfavor in recent years. This is particularly true for once-through coolant. Since its flow rate cannot be allowed to go too low—which would lead to fouling as well as excessive exit coolant temperatures, which, in turn, contribute to corrosion—it permits only limited control of pressure. Tempered coolant, which avoids these problems, is a better choice for column pressure control via coolant flow manipulation.

The signal flow diagram for pressure control via coolant flow manipulation is given in Figure 17.15. Once-through coolant and steam flow or flow-ratio control are assumed. The left-hand side of this diagram comes from Figure 17.3. Note that w_c is the rate of condensation, lbm/sec.

17.6 COLUMN ΔP CONTROL VIA HEAT TO REBOILER

The control of column ΔP by throttling steam to the reboiler was once very popular in the chemical industry, particularly for small columns. The usual practice was to run at a boilup that would give considerably more reflux than called for by design. This would usually provide a product purer than specification. In an era when it was common practice to oversize columns (low f factors, bubble-cap trays, and extra trays) and there was little concern about saving energy, this approach to control did have the advantage of usually providing a good-quality product with simple instrumentation. For today's tightly designed columns, it is technically less satisfactory, and with the rapidly rising energy costs its wastage of steam is economically unattractive. Nevertheless we still have an interest in this control technique for override purposes; an override controller is now commonly used to keep column ΔP from exceeding the maximum value specified by the column designer or determined by plant tests.

Before looking at the overall control scheme, let us discuss what is meant by the column impedance, $Z_{col}(s)$.

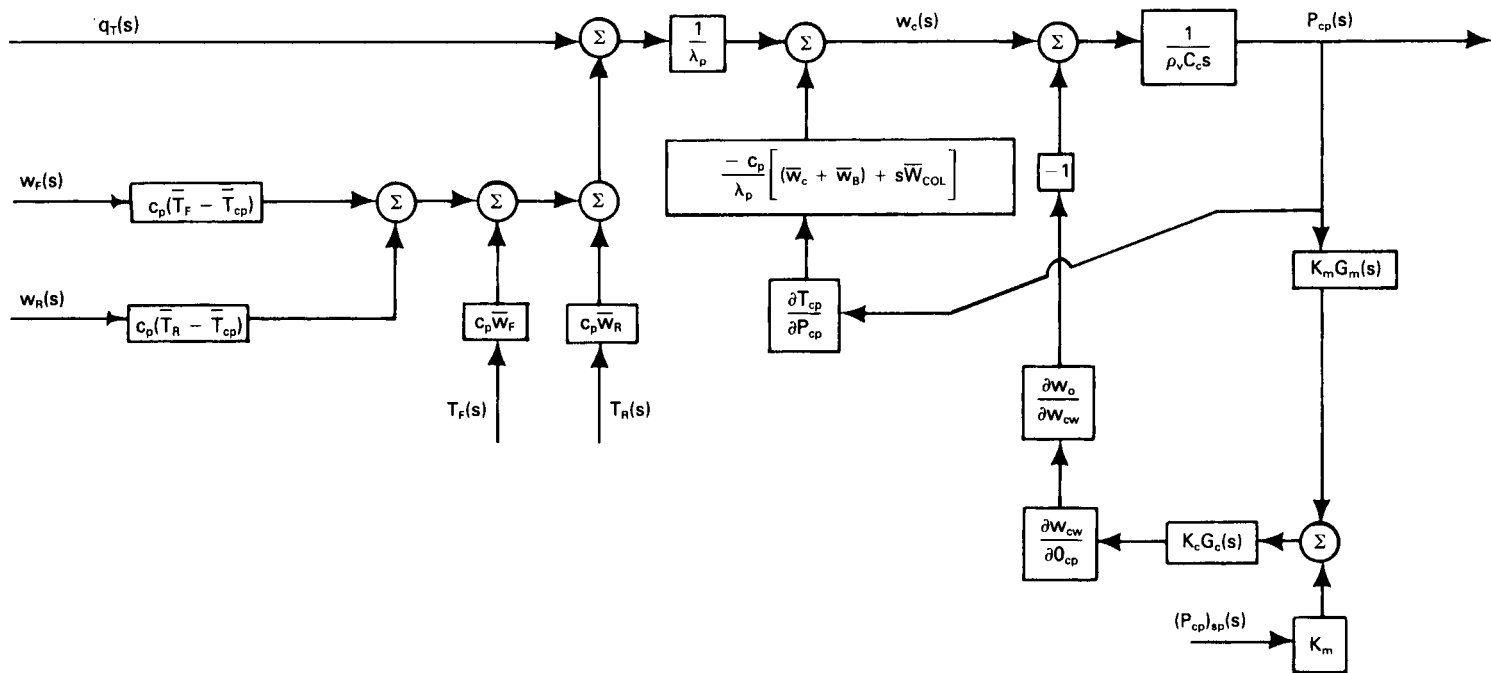


FIGURE 17.15
Signal flow diagram for column pressure
control via manipulation of condenser cooling water

Relationship Between Boilup and Column Pressure Drop

So far we have assumed that there is negligible lag in vapor flow between adjacent trays. This means that we treat the acoustic impedance Z_{col} between the bottom and top trays as a pure resistance approximately equal to

$$2 \left(\frac{\Delta \bar{P}_s}{\bar{Q}_s} + \frac{\Delta \bar{P}_r}{\bar{Q}_r} \right)$$

where the subscripts s and r refer to the stripping and rectification sections of the column. The validity of this assumption has been shown by tests run by Stanton and Bremer¹ on a 90-tray column and by the computer studies of Williams, Harnett, and Rose.²

If, however, we are interested in the high-frequency behavior of the column, then we must treat the impedance as that of an RC chain as shown in Figure 17.16. Here each RC section represents one tray, the resistance is that of the tray and layer of liquid to vapor flow, and the capacitance is the acoustic capacitance of the space between the trays. The terminal impedance, $Z_T(s)$, is simply $P_\varphi(s)/Q_c(s)$. Mathematically the entire network may be studied by the methods of transmission-line analysis.

If the individual RC sections are equivalent, or nearly so, then the impedance looking up from the reboiler is approximately:

$$\frac{P(0, s)}{Q(0, s)} \approx Z_k(s) \left[\frac{Z_T(s) + Z_k(s) \tanh nl}{Z_k(s) + Z_T(s) \tanh nl} \right] \quad (17.11)$$

where

$$Z_k(s) \approx \sqrt{\frac{R}{Cs}} \quad (17.12)$$

$$n \approx \sqrt{RCs} \quad (17.13)$$

and

$$Q(0, s) = \frac{w_v(s)}{\rho_{BU}} \quad (17.14)$$

where R and C are the resistance and capacitance, respectively, for each tray and vapor space, and l is the total number of trays. Two cases are now of primary interest.

If $Z_T = 0$, as would be true for an atmospheric column or for a column with tight overhead pressure control, then:

$$\frac{P(0, s)}{Q(0, s)} \approx Z_k(s) \tanh nl \quad (17.15)$$

$$\frac{P(0, s)}{Q(0, s)} = lR \left[1 - l^2 \frac{RC}{3} s + \frac{2}{15} l^4 R^2 C^2 s^2 - \dots \right] \quad (17.16)$$

As can be seen, the impedance becomes lR at low frequencies.

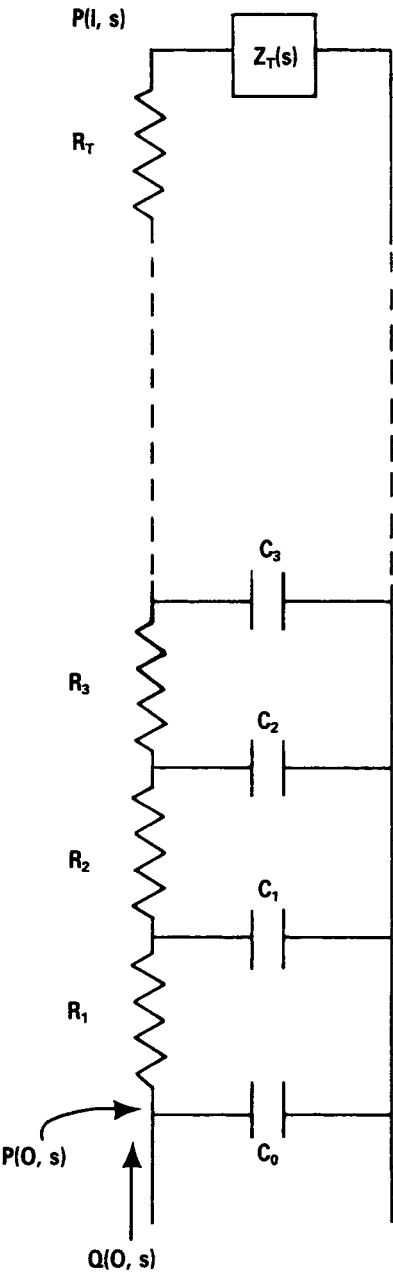


FIGURE 17.16
Equivalent network for vapor flow and pressures in column

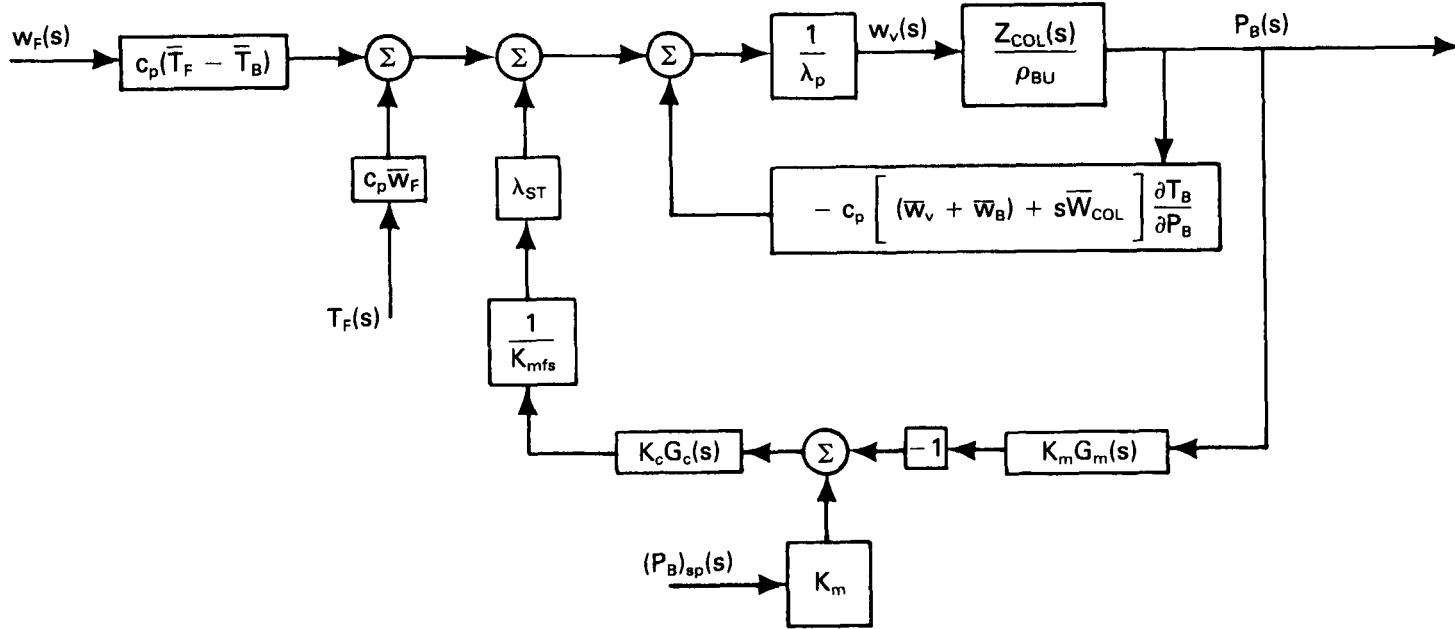


FIGURE 17.17
Column ΔP (base pressure) control cascaded to steam flow control

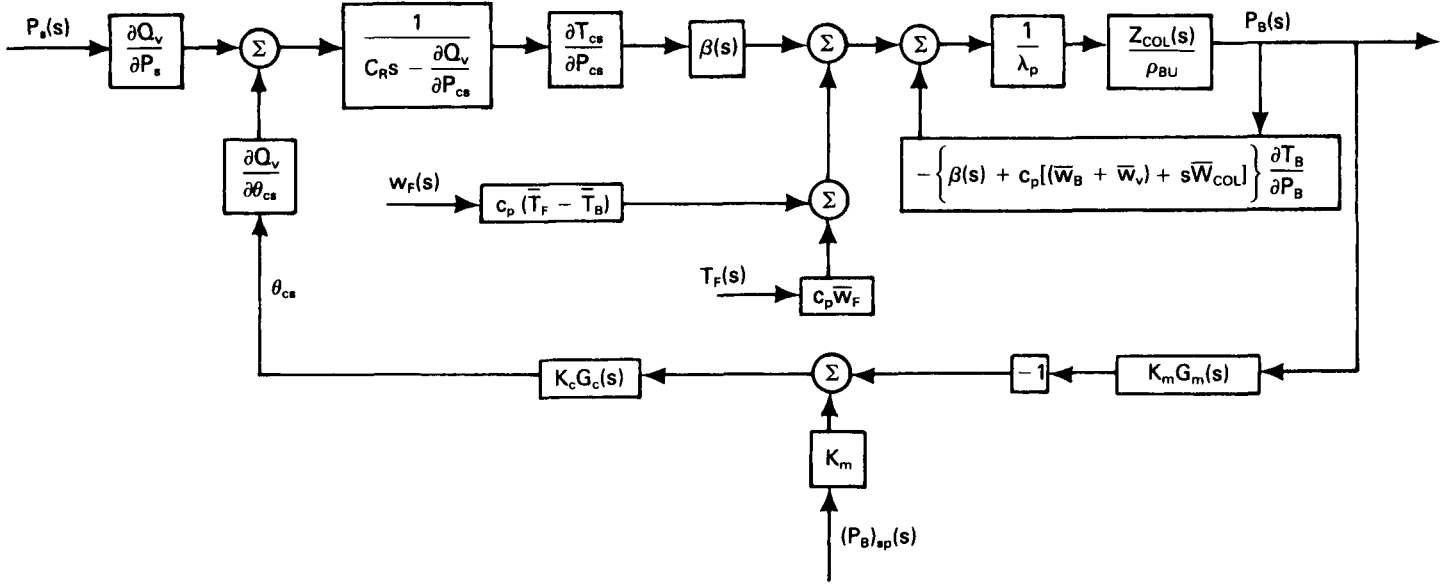


FIGURE 17.18
Column P (base pressure) control via direct manipulation of steam valve

The other case of interest is that of a tall column (many trays). Then at high frequencies:

$$\frac{P(0, s)}{Q(0, s)} \approx Z_k(s) \approx \sqrt{\frac{R}{Cs}} \quad (17.17)$$

A more detailed, more rigorous treatment of this subject will be found in Day.³

ΔP Control Cascaded to Steam Flow Control

Let us assume that column ΔP control is cascaded to steam flow control and that the latter, the secondary or slave loop, is tuned to be much faster than the primary loop. As in the case of level control cascaded to flow control, the flow loop must have a linear flow meter or an orifice meter followed by a square root extractor.

We may now simplify and rearrange Figure 15.8 and add the ΔP controller as shown in Figure 17.17. Note that for constant top pressure control, column-base pressure control and ΔP control are equivalent.

ΔP Control Via Direct Steam Valve Manipulation

If, as is usually the case, the ΔP controller is connected directly to the steam valve positioner, we need a somewhat different analysis. This calls for another rearrangement of Figure 15.8, as shown by Figure 17.18.

Note that $\beta(s)$ is defined by equation (17.3).

REFERENCES

1. Stanton, B. D., and A. Bremer, ISA Paper W-2-58, Wilmington, Del., 1958.
2. Williams, T. J., B. T. Harnett, and A. Rose, *Ind. Eng. Chem.*, 48(6): 1008–1019 (1956).
3. Day, R. L., “*Plant and Process Dynamic Characteristics*,” edited by A. J. Young, Academic Press, New York, 1957.

18

Composition Dynamics— Binary Distillation

18.1 INTRODUCTION

in Chapter 13 we looked at the material-balance dynamics of trays. We now consider composition dynamics. Since this is an extremely complex subject if discussed fully, we limit our attention to some of the more basic aspects. First, let us take up the dynamics of a single tray (not the feed tray) in the interior of the column. In the following discussion, we use a linearized treatment of the column equations for binary distillation since limited experience to date indicates that linearization does not lead to serious errors for a conventional column whose feed composition does not vary too much. For binary distillations with extremely high purity product and for multicomponent distillations where feed composition varies widely, where there is decidedly unequal molal overflow, and where the column has a number of sidestream drawoffs, more rigorous nonlinear equations should be used.

18.2 BASIC TRAY DYNAMICS

Let us visualize a single tray such as that of Figure 18.1 with an entering vapor stream from the tray below, V_{n-1} ; an entering liquid stream from the tray above, L_{n+1} ; a vapor stream leaving, V_n ; and a liquid stream leaving, L_n . The holdup on the tray is M_n moles. There are five equations of primary interest; the assumptions are those for an ideal binary distillation, equal molal overflow.

The first is the tray material balance on the more volatile component (perfect mixing assumed):

$$\frac{d}{dt}(M_n x_n) = L_{n+1} x_{n+1} - L_n x_n + V_{n-1} y_{n-1} - V_n y_n \quad (18.1)$$

Then we have the Murphree tray efficiency:

$$E = \frac{y_n - y_{n-1}}{y_n^* - y_{n-1}} \quad (18.2)$$

and the vapor–liquid equilibrium equation:

$$y_n^* = \frac{\alpha x_n}{1 + (\alpha - 1)x_n} \quad (18.3)$$

The next two equations concern the total tray material balance:

$$\frac{dM_n}{dt} = L_{n+1} - L_n + V_{n-1} - V_n \quad (18.4)$$

and

$$L_n(t) = \frac{\partial L_n}{\partial M_n} M_n(t) \quad (18.5)$$

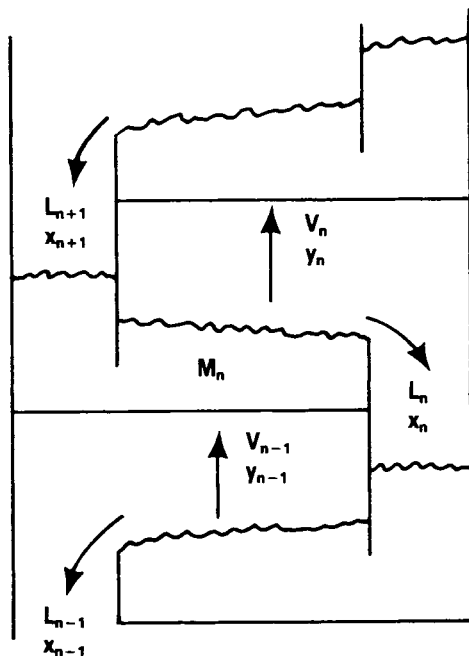


FIGURE 18.1
Flows to and from basic tray

This last equation must be used with caution and only after a test made for inverse response (see Chapter 13) shows that response to a change in boilup is neutral.

Terms in the preceding equations not already defined have the following meanings:

V and L = mols/hr

x = mol fraction low boiler (light key) in liquid

y = mol fraction low boiler (light key) in vapor

In linearizing and Laplace transforming equation (18.1), we get:

$$\begin{aligned}\bar{M}_n s \bar{x}_n(s) + \bar{x}_n s \bar{M}_n(s) &= \bar{L}_{n+1} \bar{x}_{n+1}(s) + \bar{x}_{n+1} \bar{L}_{n+1}(s) \\ &\quad - \bar{L}_n \bar{x}_n(s) - \bar{x}_n \bar{L}_n(s) + \bar{V}_{n-1} \bar{y}_{n-1}(s) \\ &\quad + \bar{y}_{n-1} \bar{V}_{n-1}(s) - \bar{V}_n \bar{y}_n(s) - \bar{y}_n \bar{V}_n(s)\end{aligned}\quad (18.6)$$

Then since the vapor composition y_n is a function of both x_n and y_{n-1} (for any tray efficiency less than unity), we can write:

$$y_n(s) = \frac{\partial y_n}{\partial x_n} x_n(s) + \frac{\partial y_n}{\partial y_{n-1}} y_{n-1}(s) \quad (18.7)$$

A rearrangement of equation (18.2) gives:

$$y_n = E y_n^* + (1 - E) y_{n-1} \quad (18.8)$$

Substituting for y_n^* from equation (18.3), we get:

$$y_n = \frac{E \alpha x_n}{1 + (\alpha - 1) x_n} + (1 - E) y_{n-1} \quad (18.9)$$

Partial differentiation of equation (18.9) leads to:

$$\frac{\partial y_n}{\partial x_n} = \frac{E \alpha}{[1 - (\alpha - 1) \bar{x}_n]^2} \quad (18.10)$$

and

$$\frac{\partial y_n}{\partial y_{n-1}} = 1 - E \quad (18.11)$$

As discussed in Chapter 17, most evidence available today indicates that there is no significant lag in vapor flow between adjacent trays, so:

$$V_n(s) \cong V_{n-1}(s) \cong V(s) \quad (18.12)$$

We can now combine equations (18.6), (18.7), (18.11), and (18.12) to give:

$$\begin{aligned}
 \left(\bar{M}_n s + \bar{L}_n + \bar{V}_n \frac{\partial y_n}{\partial x_n} \right) x_n(s) &= \bar{x}_{n+1} L_{n+1}(s) \\
 &\quad - \bar{x}_n L_n(s) + \bar{L}_{n+1} x_{n+1}(s) + \bar{V}_n E y_{n-1}(s) \\
 &\quad + (\bar{y}_{n-1} - \bar{y}_n) V(s) - \bar{x}_n s M_n(s) \\
 &\quad + (\bar{V}_{n-1} - \bar{V}_n) y_{n-1}(s)
 \end{aligned} \quad (18.13)$$

For most trays $\bar{V}_{n-1} = \bar{V}_n = \bar{V}$, and so the last term cancels out.

Next equations (18.4) and (18.5) may be transformed and combined:

$$\frac{L_n(s)}{L_{n+1}(s)} = \frac{1}{\frac{s}{\left(\frac{\partial L_n}{\partial M_n} \right)} + 1} = \frac{1}{\tau_{TR} s + 1} \quad (18.14)$$

Harriott¹ has shown that:

$$\tau_{TR} \approx \frac{A_{TR}}{\left(\frac{\partial Q}{\partial H_w} \right)} \quad (18.15)$$

where

A_{TR} = active area of the tray, which is usually
the column cross-sectional area minus
the area of two downcomers, ft²

$\frac{\partial Q}{\partial H_w}$ = change in flow over the outlet weir
per change in height over the weir,
 $\frac{\text{ft}^3/\text{sec}}{\text{ft}}$

By using the approximate form of the Francis weir formula:

$$Q \approx k H_w^{3/2} \quad (18.16)$$

we obtain:

$$\tau_{TR} \approx \frac{2}{3} \frac{A_{TR} \bar{H}_w}{\bar{Q}} \quad (18.17)$$

Typical values of τ_{TR} calculated by the authors for sieve tray columns are in the range of 2.5–8 seconds. But, as mentioned previously, the derivation of τ_{TR} is valid only if the column does not have significant inverse response.

Finally equations (18.8), (18.9), (18.13), and (18.14) may be combined (after some reduction) into the signal flow diagram of Figure 18.2 for a basic tray. The transmissions $V(s)$ and $y_n(s)$ go to the tray above; the transmissions $L_n(s)$ and $x_n(s)$ go to the tray below. In this way a signal flow diagram for any number of trays can be prepared, although the feed tray and two terminal trays have slightly modified diagrams.

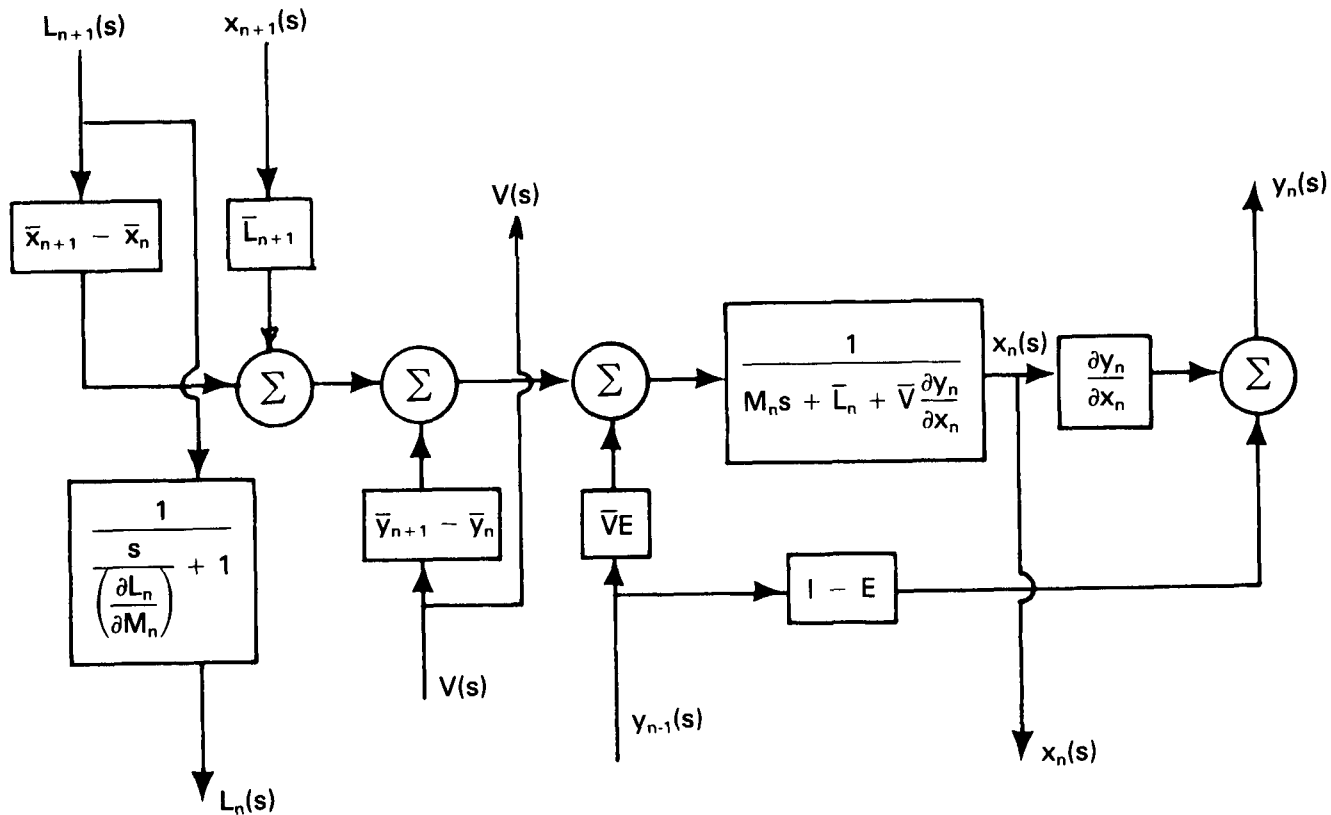


FIGURE 18.2
Signal flow diagram for basic tray

18.3 FEED TRAY DYNAMICS

The feed tray differs from the basic tray in two ways: (1) it has an additional flow, the feed flow; and (2) the thermal condition of the feed may cause the vapor and liquid rates above the feed tray to be significantly different from those below the feed tray. Let us first write an equation for the summation of steady-state flows at the feed tray:

$$F + V_{f-1} + L_{f+1} = L_f + V_f \quad (18.18)$$

Now define:

$$q = \frac{L_f - L_{f+1}}{F} \quad (18.19)$$

Substituting equation (18.19) into (18.18) and rearranging gives us:

$$V_f = F(1 - q) + V_{f-1} \quad (18.20)$$

from which, by partial differentiation, we obtain:

$$\frac{\partial V_f}{\partial F} = 1 - \bar{q} \quad (18.21)$$

$$\frac{\partial V_f}{\partial q} = -\bar{F} \quad (18.22)$$

and

$$\frac{\partial V_f}{\partial V_{f-1}} = 1 \quad (18.23)$$

We can next write a material-balance equation for the more volatile component (perfect mixing assumed):

$$\frac{d}{dt}(M_f x_f) = L_{f+1} x_{f+1} + z_F F - L_f x_f + V_{f-1} y_{f-1} - V_f y_f \quad (18.24)$$

where z_F = low boiler mole fraction in feed. When linearized and Laplace transformed, this becomes:

$$\begin{aligned} \bar{M}_f s x_f(s) + \bar{x}_f s M_f(s) &= \bar{L}_{f+1} x_{f+1}(s) + \bar{x}_{f+1} L_{f+1}(s) + \bar{F} z_F(s) + \bar{z}_F F(s) \\ &\quad - \bar{L}_f x_f(s) - \bar{x}_f L_f(s) + \bar{V}_{f-1} y_{f-1}(s) + \bar{y}_{f-1} V_{f-1}(s) \\ &\quad - \bar{V}_f y_f(s) - \bar{y}_f V_f(s) \end{aligned} \quad (18.25)$$

We can write an equation for $y_f(s)$ analogous to equation (18.7):

$$y_f(s) = \frac{\partial y_f}{\partial x_f} x_f(s) + \frac{\partial y_f}{\partial y_{f-1}} y_{f-1}(s) \quad (18.26)$$

Next the total material balance for the tray is:

$$\frac{dM_f}{dt} = F + L_{f+1} - L_f + V_{f-1} - V_f \quad (18.27)$$

which may be Laplace transformed to:

$$sM_f(s) = F(s) + L_{f+1}(s) - L_f(s) + V_{f-1}(s) - V_f(s) \quad (18.28)$$

Note that

$$L_f(s) = \frac{\partial L_f}{\partial M_f} M_f(s)$$

From equations (18.20) through (18.28) we can prepare the signal flow diagram of Figure 18.3. Note that q is a measure of the thermal condition of the feed; q is approximately the heat necessary to vaporize 1 pound mole of feed divided by the molar latent heat of vaporization of the feed. On a McCabe–Thiele diagram (see Section 2.4), the slope of the so-called q line is $q/(q - 1)$; the intercept on the 45° line is always z_F .

18.4 TOP-TRAY AND OVERHEAD SYSTEM COMPOSITION DYNAMICS

The top tray can be represented by a signal flow diagram similar to that of the basic tray (see Figure 18.2) if we make the simplifying assumption that the reflux, L_o , enters the top tray at its boiling point. Then $L_R = \bar{L}_o$. This is not usually true, but if the reflux is subcooled only a few degrees, subcooling has only a small effect on reflux enthalpy. Further, the reflux temperature is sometimes controlled, so the reflux enthalpy does not change significantly. For those cases where it is not practical to control reflux temperature, it is usually possible to control internal reflux rate. This we are doing more frequently today. Ignoring subcooling, therefore, in most cases leads to small errors in the calculated values of static gains and top-tray mixing time constant.

We assume further that the vapor from the top tray is totally condensed. We can then write the following transfer function relating the vapor composition y_T and reflux composition x_R :

$$\frac{x_R(s)}{y_T(s)} = \frac{e^{-(a_1 + a_2)s}}{\tau_D s + 1} \quad (18.29)$$

where

a_1 = vapor-flow transport delay from top tray to condensate receiver, minutes

τ_D = mixing time constant of well-mixed condensate receiver, minutes

a_2 = liquid-flow transport delay from condensate receiver back to top tray, minutes

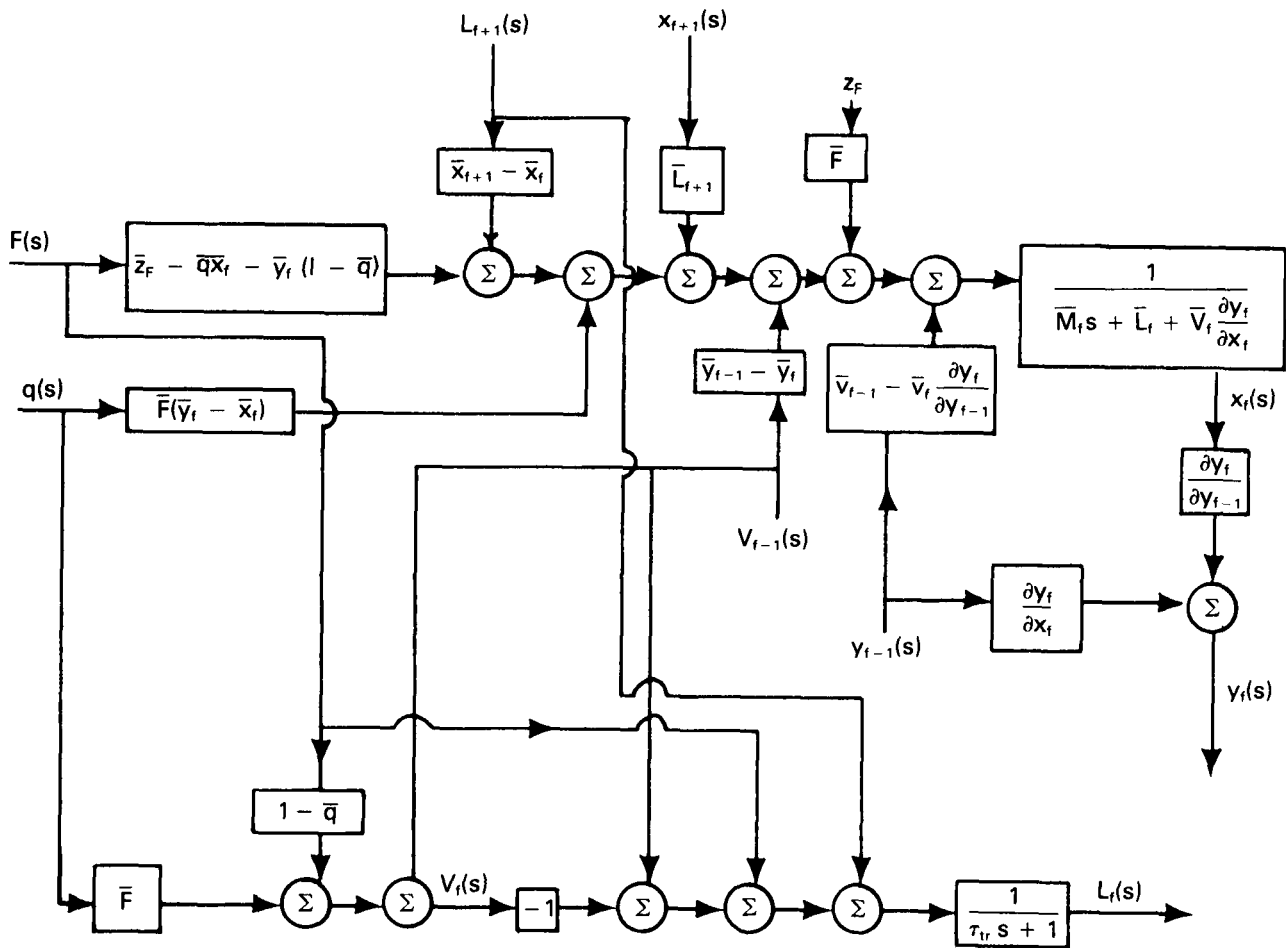


FIGURE 18.3
Signal flow diagram for feed tray

The signal flow diagram of Figure 18.4 may now be prepared. Note that a composition control loop is also shown. Note that y_{-1} is the mol fraction of low boiler in the vapor leaving the tray below the top tray.

The term $L_R(s)/\theta_c(s)$ relates reflux flow to controller output signal; for the moment there are no implications as to whether the manipulation is simply a valve, the set point to the secondary flow controller, or the ratio in a ratio-control system.

By collapsing the overhead composition loop and the two transmissions for $y_{-1}(s)$, we are led to the signal flow diagram of Figure 18.5. Note that:

$$\beta(s) = \frac{1}{1 - \frac{\bar{L}_R \frac{\partial y_T}{\partial x_T} e^{-(a_1 + a_2)s}}{(\tau_D s + 1) \left(\bar{M}s + \bar{L}_R + \bar{V} \frac{\partial y_T}{\partial x_T} \right)}} \quad (18.30)$$

Rippin and Lamb² point out that if we provide a perfect feedforward control system to adjust L_R so that feed changes in flow and composition do not change y_T , then the transmission $x_R(s)$ is broken and there is no feedback of $x_R(s)$ down the column. It is possible to accomplish the same thing by making $1/\tau_D$ sufficiently smaller than the resonant frequency of the closed-loop composition-control system. This has been shown experimentally by Aikman³ to be true of a plant column. The mathematical explanation is simple.

If dead time is negligible, then we may write:

$$\beta(s) = \frac{1}{1 - \frac{K}{(\tau_D s + 1)(\tau_T s + 1)}} \quad (18.31)$$

where

$$K = \frac{\bar{L}_R \frac{\partial y_T}{\partial x_T}}{\bar{L}_R + \bar{V} \frac{\partial y_T}{\partial x_T}} \quad (18.32)$$

and

$$\tau_T = \frac{\bar{M}}{\bar{L}_R + \bar{V} \frac{\partial y_T}{\partial x_T}} \quad (18.33)$$

This is the mixing time constant of the top tray. Then:

$$\beta(s) = \frac{(\tau_D s + 1)(\tau_T s + 1)}{\frac{\tau_D \tau_T}{1 - K} s^2 + \frac{\tau_D + \tau_T}{1 - K} s + 1} \left(\frac{1}{1 - K} \right) \quad (18.34)$$

$$= \frac{(\tau_D s + 1)(\tau_T s + 1)}{\tau_Q^2 s^2 + 2 \zeta \tau_Q s + 1} \left(\frac{1}{1 - K} \right) \quad (18.35)$$

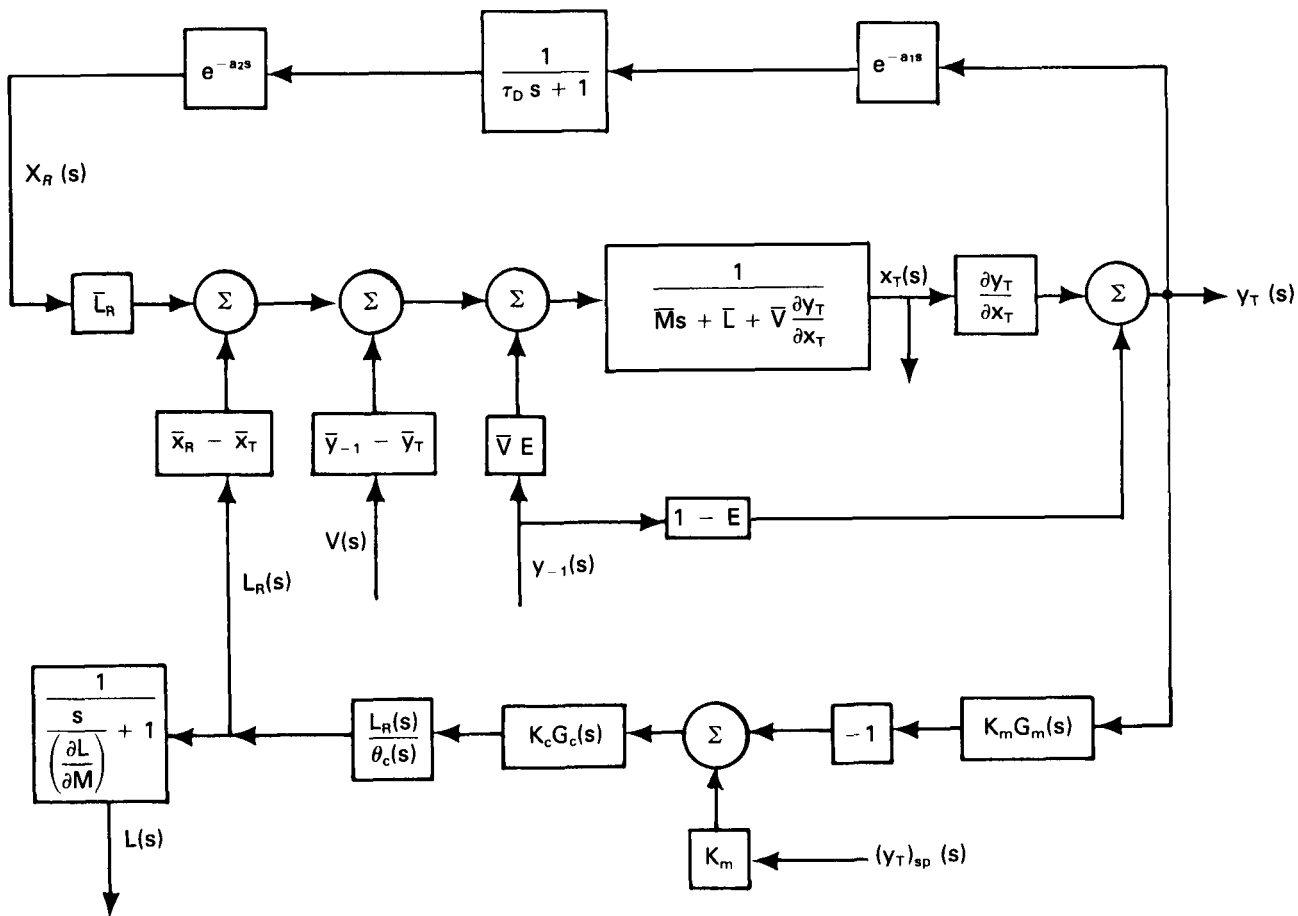


FIGURE 18.4
Signal flow diagram for top tray and overhead system

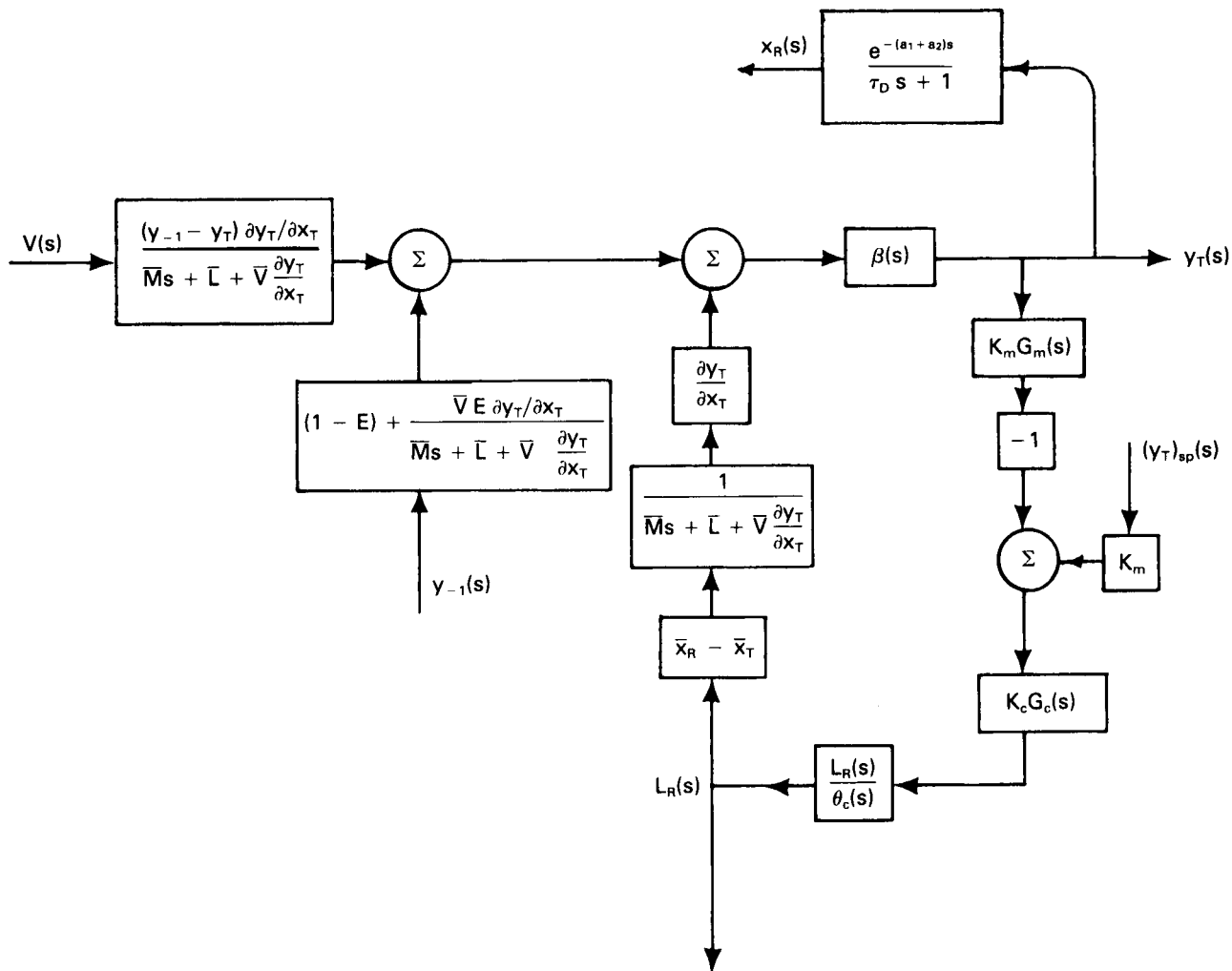


FIGURE 18.5
Partly reduced signal flow diagram for top tray and overhead system

where

$$\tau_Q = \sqrt{\frac{\tau_D \tau_T}{1 - K}} \quad (18.36)$$

and

$$\zeta = \frac{\tau_D + \tau_T}{2 \sqrt{\frac{\tau_D \tau_T}{1 - K}}} \times \frac{1}{1 - K} \quad (18.37)$$

If $\tau_D \gg \tau_T$, then:

$$\zeta = \frac{1}{2} \sqrt{\frac{\tau_D}{(1 - K) \tau_T}} \gg 1 \quad (18.38)$$

The denominator of equation (18.35) may now be factored into two first-order lags such that the larger is:

$$\tau_a \approx 2 \zeta \tau_Q = \frac{\tau_D}{1 - K} \quad (18.39)$$

and the smaller is:

$$\tau_b \approx \frac{\tau_Q}{2 \zeta} \approx \tau_T \quad (18.40)$$

Substituting back into equation (18.35), we get:

$$\beta(s) = \frac{(\tau_D s + 1)(\tau_T s + 1)}{\left(\frac{\tau_D}{1 - K} s + 1\right)(\tau_T s + 1)} \left(\frac{1}{1 - K}\right) \quad (18.41)$$

$$= \frac{\tau_D s + 1}{\frac{\tau_D}{1 - K} s + 1} \left(\frac{1}{1 - K}\right) \quad (18.42)$$

At frequencies below $(1 - K)/\tau_D$, $[\beta(j\omega)]$ has a gain of $1/(1 - K)$ while at frequencies above $1/\tau_D$ it has a gain of unity. Therefore, if the resonant frequency of the closed-loop composition control system is well above $1/\tau_D$, then there is in effect no transmission of x_R . As a practical design consideration, the low-frequency phase-shift bulge of a proportional-reset controller should be so set that it does not coincide with the phase shift bulge in $\beta(j\omega)$. Otherwise conditional stability might occur if $1/(1 - K)$ is large enough.

Although we have indicated two methods of breaking the $x_R(s)$ transmission, thereby simplifying composition control system design and improving the control of $y_T(s)$, it should be noted that the first procedure—feedforward control—is more desirable than the use of a large condensate receiver. Actually the control of top-tray vapor composition is not as important as the control of composition of condensate to the next step. It can be shown that blending is much more effective if done outside the composition control loop.

Finally it should be noted from equation (18.30) that it is quite important to keep the dead time in the overhead system to a minimum. This means that transport times in the vapor and reflux lines should be as small as possible. The condensate receiver, if unagitated, should have as small a holdup as possible. See, however, the discussion in Chapter 3.

18.5 REBOILER AND COLUMN-BASE COMPOSITION DYNAMICS

If the reboiler is either a well-agitated kettle type or a thermosyphon type with a high recirculation rate, the reboiler/column-base composition dynamics are essentially those of a simplified basic tray (see Figure 18.1). The chief difference is that the reboiler has no entering vapor flow from a lower tray. Usually, too, the reboiler has considerably more holdup than a typical tray. The holdup, M_B , is that of the reboiler itself plus that of the column base.

We therefore may prepare the signal flow diagram of Figure 18.6. L_1 and x_1 are the liquid flow and its composition from the lowest tray. A composition control system is shown in which $V(s)/\theta_c(s)$ is the transfer function relating vapor flow to composition controller output. It is assumed that steam flow is controlled by a steam flow controller, but for the moment there are no implications as to whether the steam flow controller set-point signal comes directly from the composition controller or from a steam-to-feed ratio controller that is reset by the composition controller.

18.6 INVERSE RESPONSE

The phenomenon of inverse response was apparently first noted by Rijnsdorp.⁴ As mentioned in Chapter 13, it occurs in some columns when an *increase* in vapor flow causes a *momentary increase* in liquid flow down the column. This is due to a decrease in foam density on the trays and a consequent momentary overflow into the downcomer.

As may be seen from Figure 18.2–18.4 and 18.6, an increase of liquid flow onto a tray increases low boiler concentration. Therefore, although in the long run a vapor flow *increase* will *decrease* low boiler concentration, in the short run it *increases* it. This momentary change of concentration in the wrong direction gives rise to the term “inverse response.”

The significance of inverse response in the design of feedforward compensators and feedback controllers for composition is discussed later. It is the subject of a paper by Luyben.⁵

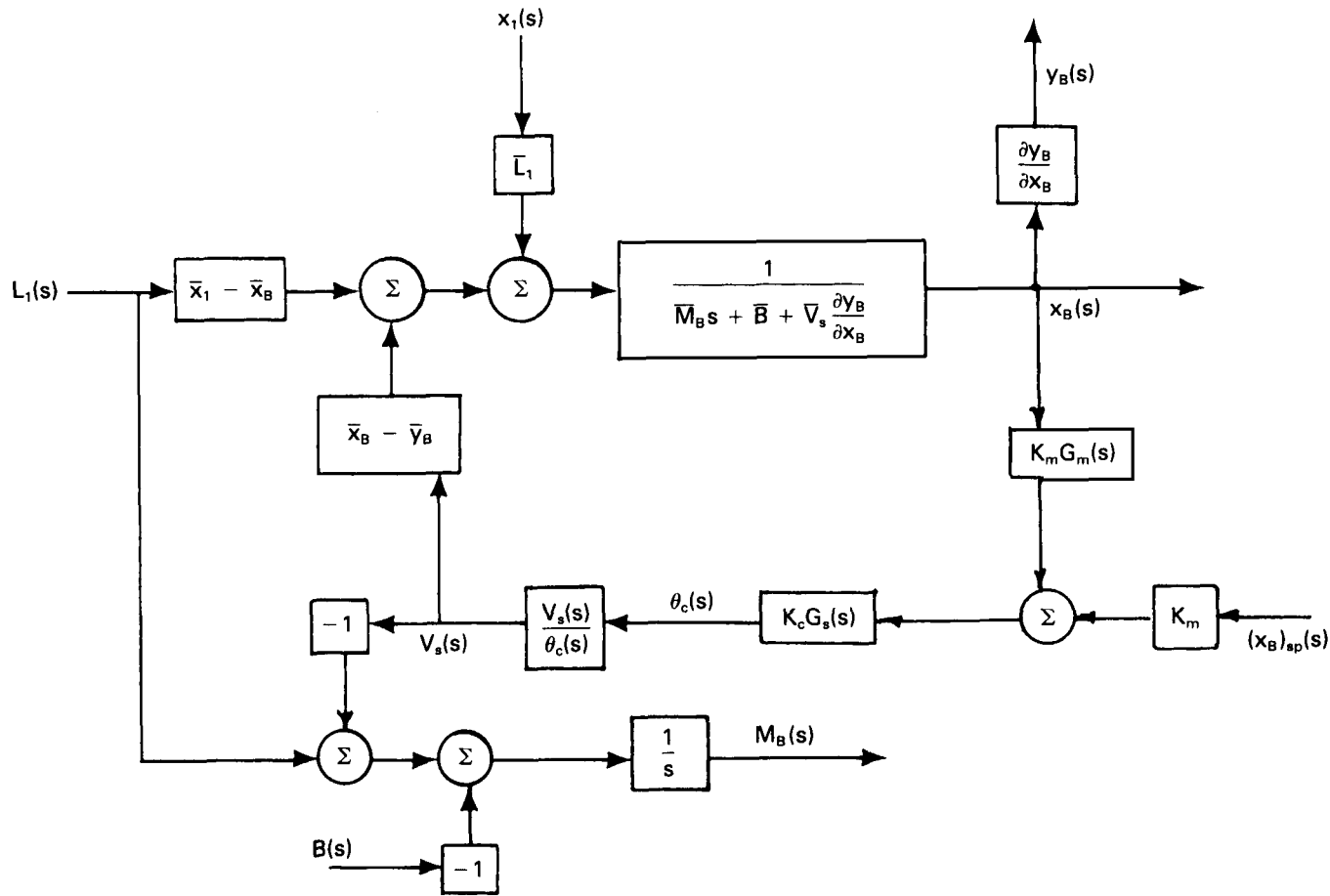


FIGURE 18.6
Signal flow diagram for reboiler composition dynamics

18.7 OVERALL COMPOSITION DYNAMICS

In previous sections we examined in some detail the composition dynamics of the feed tray, overhead system, and bottom system. This was primarily to see how feed disturbances enter the column and how the design and arrangement of auxiliaries can affect column dynamics. From the material previously presented, one can combine the various equations affecting column composition dynamics and thereby calculate overall column behavior. Teager⁶ and Campbell⁷ suggest that this combining might be done in two ways: (1) by connecting the signal flow diagrams of the individual trays and auxiliaries, and (2) by preparing all of the column equations in matrix form.

The signal flow diagram approach to deriving individual transfer functions is exceedingly tedious even for a short column, and even when the shorthand notation of signal flow graphs is used. For this reason more recent studies have usually been based on matrix methods, the "stepping" technique, or a third method—simulation on an analog or digital computer.⁸ A small amount of work has been done with hybrid computers.

Binary Distillation—Rippin and Lamb Model

For a binary distillation column, convenient, compact representation of overall column composition dynamics in two equations has been suggested by Rippin and Lamb:²

$$\begin{aligned}
 y_T(s) = & \left[\frac{y_T(s)}{z_F(s)} \right]_{OL} z_F(s) + \left[\frac{y_T(s)}{F(s)} \right]_{OL} F(s) \\
 & + \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} L_R(s) + \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} V_s(s) \\
 & + \left[\frac{y_T(s)}{q(s)} \right]_{OL} q(s) + \left[\frac{y_T(s)}{p(s)} \right]_{OL} p(s)
 \end{aligned} \quad (18.43)$$

where p is the top pressure in psia and

$$\begin{aligned}
 x_B(s) = & \left[\frac{x_B(s)}{z_F(s)} \right]_{OL} z_F(s) + \left[\frac{x_B(s)}{F(s)} \right]_{OL} F(s) + \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} L_R(s) \\
 & + \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} V_s(s) + \left[\frac{x_B(s)}{q(s)} \right]_{OL} q(s) + \left[\frac{x_B(s)}{p(s)} \right]_{OL} p(s)
 \end{aligned} \quad (18.44)$$

where the individual transfer functions are determined by one of the methods indicated above. Forcing functions in the original Rippin and Lamb equations were limited to L_R , V_s , z_F , and F ; the authors have added terms for p and q . Symbolism such as

$$\left[\frac{y_T(s)}{z_F(s)} \right]_{OL}$$

means the open-loop transfer function of y_T with respect to z_F . Equations (18.43) and (18.44) may now be represented in signal flow diagram form as shown in Figure 18.7.

Rippin and Lamb developed a stepping procedure, later extended by Luyben and others,⁹⁻¹⁴ for computing these transfer functions in the frequency domain with a digital computer. Applying asymptote techniques to the Bode plots, they determined approximate Laplace transforms. Since these transfer functions were mostly of relatively low order—second to fourth order—they could be simulated on an analog computer with much less hardware than would be necessary with more conventional distillation-column simulations. Feedback control for top and bottom composition could then be designed with ease. The advent, however, of more powerful and less expensive digital computers has shifted the emphasis to digital simulation. More commonly today the individual tray differential equations are combined and solved in the time domain.

A comparison of the computational efficiency of stepping and matrix-inversion techniques has been carried out by Shunta and Luyben.¹⁵ For large columns in particular, it is shown that the former is much faster.

It should be noted that some of the work of modeling column composition dynamics has been concerned with multicomponent rather than binary separations.¹²⁻¹⁴

It should also be noted that the open-loop transfer functions given in equations (18.43) and (18.44) are valid, strictly speaking, for only one combination of holdup volumes in the condensate receiver and the column base. Studies to date, however, suggest that if these holdups are no larger than the total holdup on the trays, then the open-loop transfer functions are relatively insensitive to variations in terminal volumes. This topic definitely needs more study. In the meantime we should determine terminal volumes by material-balance and protective-control calculations before calculating column composition dynamics.

Other Models

Two other models for binary columns are those proposed by Wahl and Harriott¹⁶ and by Waller.¹⁷ These are similar in that they are both based on circulation rates and liquid holdup in the column. Low-order transfer functions are calculated from steady-state data. The two models give essentially the same results except that the Waller method leads to higher order—and intuitively more accurate—transfer functions. For columns with moderate relative volatilities—say 2 to 5—and for terminal purities not greater than 98 to 99 percent, predicted behavior checks with that of more rigorous models. But for high-purity, nonideal separations, accuracy falls off. More work needs to be done to check the range of validity of these models.

A comprehensive review of the literature of distillation dynamics and control through about 1974 is given by Rademaker, Rijnsdorp, and Maarleveld.¹⁸ Tolliver and Waggoner¹⁹ have published an exhaustive review of more recent additions to the literature.

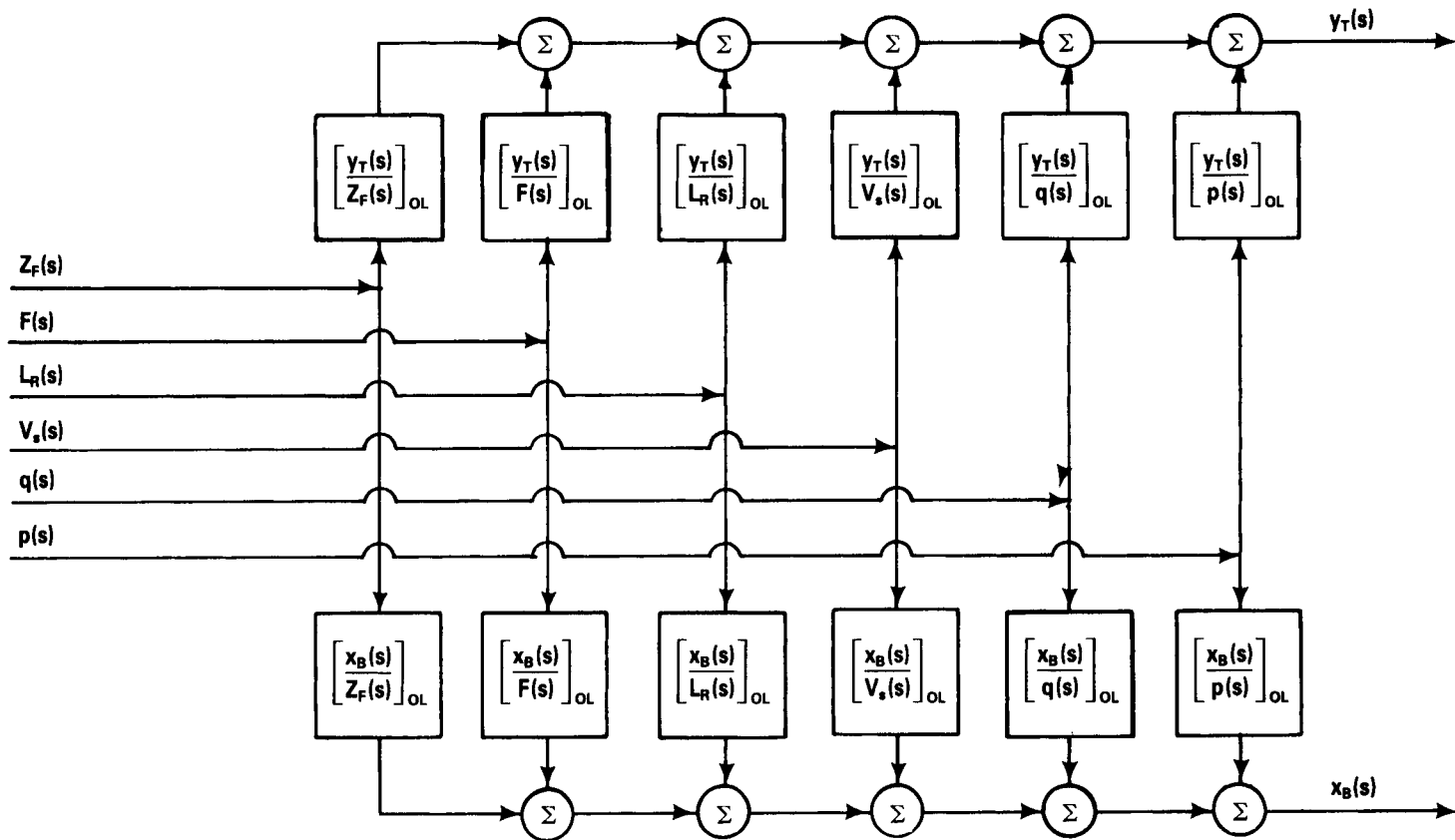


FIGURE 18.7
Signal flow diagram for Rippl-Lamb model for binary distillation column dynamics

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19

Calculation of Steady-State Gains

19.1 INTRODUCTION

most composition control schemes for a binary column involve manipulating either reflux or distillate to control top composition, and either boilup or bottom product to control bottom composition. Changes in either of the two top manipulative variables will affect not only top composition, but bottom composition as well. Correspondingly, changes in either bottom-product flow or boilup will affect both bottom and top compositions. As a minimum, therefore, for feedback control purposes we are interested in two composition gains at each end of the column.

The calculations to predict how a column already designed or built will operate are intrinsically more difficult than are column design calculations. As pointed out in a paper by Douglas and Seemann,¹ "The number of plates required to achieve a given separation at a specified reflux ratio can be calculated directly, but is implicit in terms of the control variables, i.e., the reflux rate required to achieve a given separation if the number of plates is specified must be obtained by trial-and-error procedures." An appreciation of the labor required may be obtained by a perusal of papers such as those of Utti,² and Bauer and Orr,³ who made trial-and-error use of McCabe–Thiele diagramming. More recently Wood⁴ used a digital computer for trial-and-error calculations via the tray-to-tray method, while Buckley, Cox, and Luyben⁵ performed similar calculations on a programmable calculator.

Other workers in the field seem primarily interested in such calculations to determine the economic penalty of either overrefluxing (via excess boilup) or of specifying product purities better than needed. To obtain column gains, we need go only one step further. For example, it is only necessary to make a number of top and bottom composition calculations for different reflux ratios, plot the points of top and bottom purities versus reflux ratio, and draw curves

through the points. Then the slope of either curve is the “gain” of top or bottom composition as a function of reflux ratio.

Some workers have used the Smoker analytic method rather than the tray-to-tray method. The former has the disadvantage of requiring the assumption of constant relative volatility but has the advantage of being much faster when programmable calculators or small computers are used. However, the tray-to-tray method easily handles relative volatility as a function of composition, and reads out any or all tray compositions. We feel that the advantages of the tray-to-tray method outweigh the disadvantage of longer run times.

Most industrial columns are not binary and do not have constant relative volatility from top to bottom. Two fundamental problems therefore confront the engineer at the outset:

1. How to make a reasonable binary approximation of a multicomponent column. Such an approximation greatly reduces computation and time.
2. How to derive a relationship between liquid composition on each tray and relative volatility.

As is shown here, the second problem must be dealt with first. The following procedure is suggested:

1. Pick a light key and a heavy key—or a low boiler and a high boiler.
2. Using printouts of multicomponent design calculations, calculate mol fraction low boiler, x_{LB} , on a binary basis and relative volatility at each tray.
3. Use curve-fitting techniques to relate x_{LB} and relative volatility at each tray.
4. Go through binary design calculations to check the total number of theoretical trays and feed-tray location. If the match with the multicomponent calculations is not adequate, go to step 3 and use a higher order function for curve fitting.

The basic approach then consists of defining a base case, usually the design case, to firm up feed-tray location (number of trays above and below the feed tray), reflux ratio, and boilup ratio. With these in hand, we may calculate the effect of changing reflux ratio, and so forth.

19.2 DESIGN PROCEDURE

Since we will use the tray-to-tray method, it is appropriate to begin by stating the required assumptions:

1. Simple column with a single feed and top- and bottom-product drawoffs.
2. Binary or pseudobinary separation. As shown by Strangio and Treybal,⁶ surprisingly good accuracy can be obtained in many cases by approximating a multicomponent system as a binary system. Sometimes a multicomponent system

can be modeled as one binary in the bottom section and as a different binary in the upper section.

3. Equimolar overflow.
4. One top product that is all liquid.
5. One bottom product that is all liquid.
6. Feed that may be single-phase vapor or liquid, or may be mixed phase in nature.
7. Partial reboiler and total condenser.

In the first programmable calculator programs we developed, three choices were available for relative volatility: (1) constant, (2) a linear relationship between α and liquid composition x , and (3) a quadratic relationship between relative volatility and liquid composition on a tray. But the first time we tried these on a real column for MeOH-H₂O, it proved inadequate. We now have an HP-41C program that fits a quadratic to each of six segments of a column to relate relative volatility and tray number.

The design program (see also Chapter 2) starts with data provided by the column designer or plant. This includes feed flow rate, composition, and enthalpy or q . It also includes top- and bottom-product compositions and reflux ratio, L_R/D . Running a design program is not absolutely necessary for our purposes, but it is one way of initializing subsequent programs and of checking assumptions. The following steps are involved.

1. Find:

$$\beta = \frac{V_s}{B} \quad (19.1)$$

$$\beta = \frac{R(z_F - x_B) + q(x_D - x_B) - (x_D - z_F)}{(x_D - z_F)} \quad (19.2)$$

2. Find:

$$y_i = \frac{(\beta + 1)x_D + Rx_B}{(\beta + 1) + R} \quad (19.3)$$

This is the y corresponding to the intercept of the operating and q lines. For optimum design it is also the composition of vapor above the feed tray.

3. The tray-to-tray calculations come next. If trays are numbered from the top down, it is customary to start at the top. If trays are numbered from the bottom up, the base is the usual starting point. We will follow the second practice.

- 3.1 Starting at the column base and working upward, find:

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n} \quad (19.4)$$

This is the equation defining the vapor-liquid equilibrium relationship.

It is the x versus y curve for a McCabe–Thiele diagram. Note that the first x is x_B .

3.2 Find:

$$x_{n+1} = \frac{\beta y_n + x_B}{\beta + 1} \quad (19.5)$$

This is the material-balance equation and its initial application gives x_1 , liquid composition on the bottom tray.

3.3 Repeat 3.1, using x_1 to calculate y_1 . Then use y_1 in 3.2 to calculate x_2 , and so on until $y \geq y_i$. The number of times that we perform 3.1 [equation (19.4)] is N_s , the number of stages there are in the stripping section, including feed tray and the reboiler (assuming there is one). The number of times that we perform 3.2 [equation (19.5)] is one less, and therefore represents the number of theoretical trays, including the feed tray.

4. Starting with the tray above the feed tray, we switch operating lines.

4.1 Find:

$$x_{n+1} = \frac{(R + 1)y_n - x_D}{R} \quad (19.6)$$

This is the material-balance equation for the rectification section. The first x calculated is that of the first tray above the feed tray; the value of y used is that of the feed tray.

4.2 Find:

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n} \quad (19.7)$$

4.3 Repeat 4.1 and 4.2 until $y \geq x_D$ (total condenser assumed). The number of times 4.1 and 4.2 have been repeated is N_R , the number of stages in the rectification section. Since the condenser is assumed to be a total condenser, it does not count as a stage. A partial condenser would count as a stage.

19.3 EXACT R PROCEDURE

The “design” program usually leads to a design that produces top-product purity slightly better than specification for an integral number of trays (if we start at the top instead of the bottom, it produces bottom-product purity slightly better than specification). What we need to do next is to find a slightly different, “exact” value of R that causes the column to produce top and bottom purities that exactly match specifications. This may be determined by the following procedure, designated “Type A”:

1. Assume a new value of R slightly smaller than that used for design.
2. Calculate β from equation (19.2).

3. Starting at the column base and working upward, find:

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n} \quad (19.4)$$

$$x_{n+1} = \frac{\beta y_n + x_B}{\beta + 1} \quad (19.5)$$

Repeat until the number of times equation (19.4) has been used equals N_f .

4. Use the y thus calculated for the feed tray. Then switch operating lines and find:

$$x_{n+1} = \frac{(R + 1)y_n - x_D}{R} \quad (19.6)$$

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n} \quad (19.7)$$

If the calculated $x_{n+1} < x_n$, assume a slightly larger value of R and start again at step 1. As indicated by Figure 19.1, we must be careful to provide program logic to prevent “taking off” to the left of the equilibrium curve.

5. Repeat equations (19.6) and (19.7) until they have been used $N_T - N_S = N_R$ times.

6. Compare last $y = y_T$ with specified x_D . If y_T does not match within a preselected tolerance, go back to step 1, assume a new value of R , and repeat steps 2–6.

A number of convergence procedures are available for step 6. We have, for example, used a simple “proportional controller,” that is, we multiply the “error” $y_T - x_D$ by a constant with the proper polarity for negative feedback. This requires some trial and error to achieve a balance between oscillation and excessively slow response. A better method is a modified interval-halving technique that provides moderately rapid convergence without instability.

Originally we used 0.0002 as a tolerance: $[y_T - x_D]$. This proved to be too large for some purposes so we switched to 10^{-7} .

19.4 COLUMN OPERATION PROCEDURE

Many different modes of column operation and control can be visualized, but for preparing programs, we must keep in mind at least four needs:

1. Column gains for feedback control. Here, for example, we need to know the change in top composition due to a change in distillate or reflux flow with other variables held constant.

2. Column terminal composition sensitivity to changes in feed rate, feed enthalpy, feed composition, and column pressure (usually defined at column

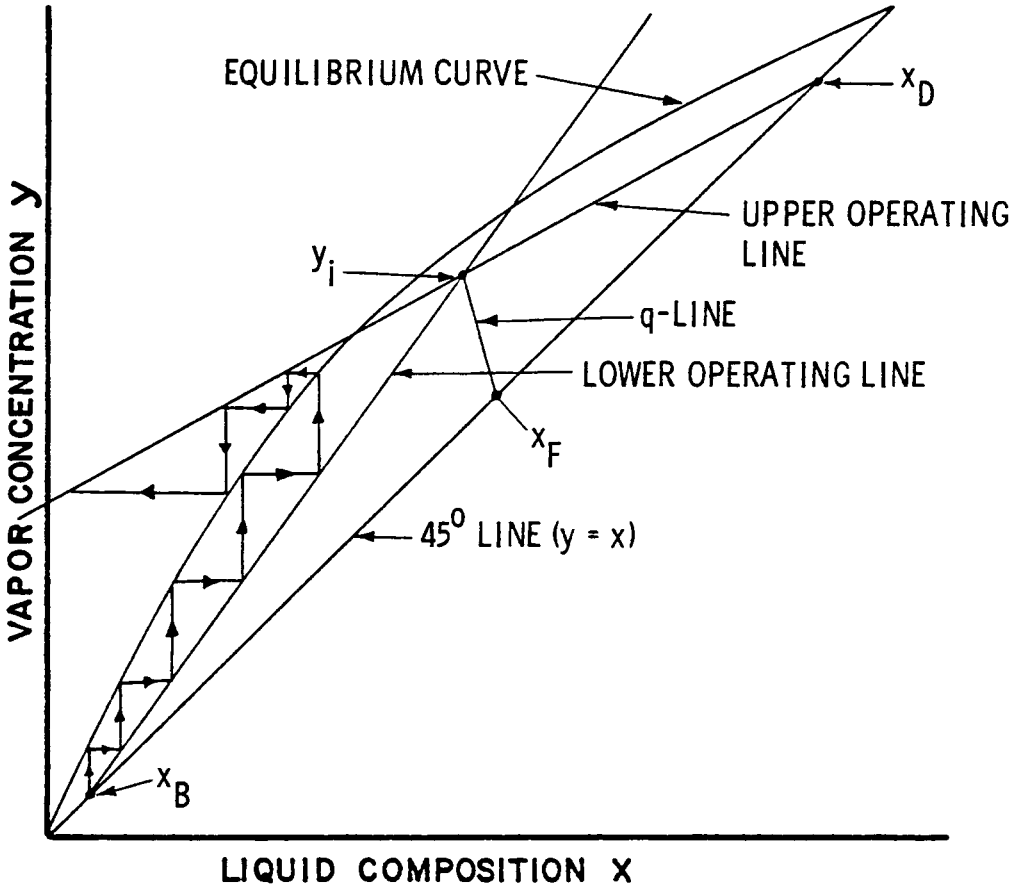


FIGURE 19.1
 Effect on calculation of rectifying section when:
 A. Guess for R is too small, or
 B. Guess for x_D is too close to 1.00000

overhead). These sensitivities often give good clues to the need for feedforward compensation or for terminal composition feedback control.

3. Column gains for feedforward compensation. For example, if feed composition changes, what changes in boilup and reflux do we need to hold terminal compositions constant? Steady-state accuracy is required.

4. Ability to estimate the cost penalty of producing excessively pure product by excessive boilup and reflux. Shinskey⁷ and Douglas and Seemann¹ have been particularly interested in this. Again, steady-state accuracy is required.

As it turns out, for most of these concerns, we use one of two types of programs: (1) Column terminal conditions are fixed at new, different values from those of design, and reflux is varied by trial and error until the material-balance equation and tray-to-tray calculations converge, or (2) with fixed reflux or boilup, we change a terminal condition such as distillate flow and find the corresponding change in terminal compositions. The first of these we have labeled "Type A" and the second we call "Type B."

Gains for Feedback Control

Case 1

Consider Figure 19.2 where top-product flow is set by flow control, reflux flow is set by condensate receiver level control, boilup is fixed by flow control of steam or other heating medium, and bottom-product flow is determined by column-base level control. As shown by the dotted line, we wish eventually to control column top composition by manipulating distillate flow. Let us assume that feed rate, feed composition, feed enthalpy, and boilup are fixed and that we wish to find the changes (i.e., "gains") of top and bottom compositions in response to a change in D , the top-product rate.

The starting point is a set of "prep" equations that reflect the desired change in distillate, ΔD . The new distillate flow is:

$$D' = D + \Delta D$$

and

$$B' = B - \Delta D \quad (\text{since } F \text{ is fixed at } 1.00)$$

$$\beta' = V_s/B'$$

$$L'_R = L_R - \Delta D$$

$$R' = L'_R/D'$$

1. With these equations in hand, assume a new value of $x_D = x'_D$ and calculate a new x'_B :

$$x'_B = z_F + \frac{D'}{B'}(z_F - x'_D)$$

2. Calculate:

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n}$$

starting at the column base and working up.

3. Calculate:

$$x_{n+1} = \frac{\beta' y_n + x'_B}{\beta' + 1}$$

Continue calculating y and x alternately until the number of times y has been calculated equals N_s .

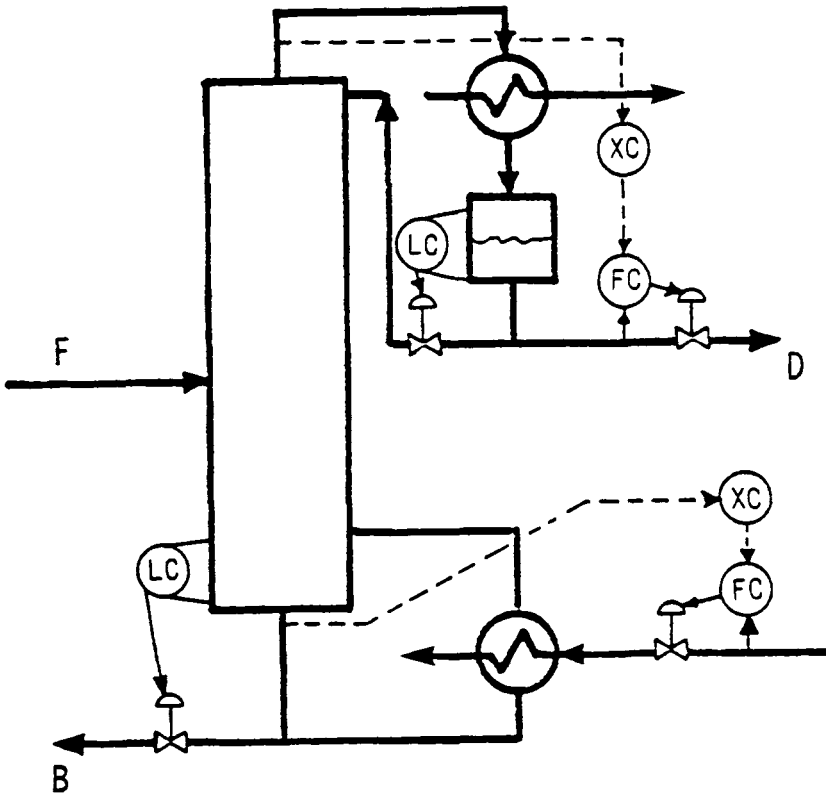


FIGURE 19.2

Reflux via reflux drum level control; bottom product via base level control

4. Switch to rectification operating line.

4.1 Calculate:

$$x_{n+1} = \frac{(R' + 1)y_n - x'_D}{R'}$$

and

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n}$$

4.2 If the first calculated x shows that $x_k < x_{k-1}$ (Figure 19.1), assume a new and larger value of x_B and resume at step 4. Continue until $N = N_T - N_S = N_R$.

5. Determine when the calculation is complete, that is, when the estimated $x'_D \leq y$ at the top tray. If $[y_T - x_D]$ is greater than the chosen tolerance, choose new x'_B and repeat steps 1–5.

Note that changing B is equivalent for this case to changing D . We will term the procedure for steps 1–5 the Type B procedure in contrast to the Type A procedure.

The reader may have observed that although we need only the change in x_D for a change in D to design the overhead composition control system, we have also calculated the change in x_B in response to a change in D . This second gain is an interaction term. Note that we call the preceding Type B.

Since the original design calculations are unconverged, it is necessary, when the Type B program is to be used, first to obtain converged values of x_B and x_D for the specified reflux and boilup.

Case 2

If we desire to control bottom composition by changing boilup as shown on Figure 19.2, we would start by assuming that both feed and top-product rates are fixed. This means that bottom product is also fixed. The “prep” equations for ultimately finding new base composition, x'_B , and new top composition, x'_D , are therefore:

$$V'_s = V_s + \Delta V_s$$

$$L'_R = L_R + \Delta V_s$$

$$\beta' = V'_s/B$$

$$R' = L'_R/D$$

Then use the Type B routine.

To control base composition via boilup, we need the gain of x_B with respect to V_s . The second gain, change in x_D with respect to V_s , is another interaction term. The magnitudes of the interaction terms give clues to the need for decouplers.

Case 3

Another commonly encountered control scheme is that of Figure 19.3. This is similar to that of Figure 19.2 except that overhead condensate receiver level is controlled by throttling distillate, while reflux is flow controlled, perhaps eventually cascaded from top composition control.

If feed and boilup are fixed at design conditions, top and bottom composition x'_D, x'_B response to reflux changes may be found by starting with the following "prep" equations:

$$L'_R = L_R + \Delta L_R$$

$$B' = B + \Delta L_R$$

$$\beta' = V_s/B'$$

$$D' = D - \Delta L_R$$

$$R' = L'_R/D'$$

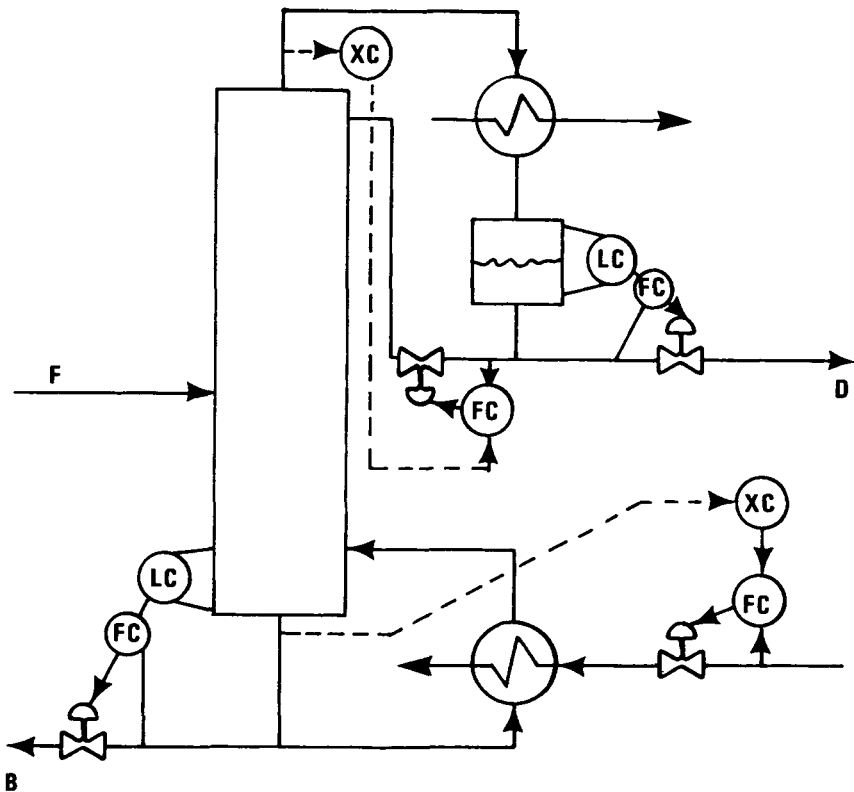


FIGURE 19.3

Distillate via reflux drum level control; bottom product via base level control

The Type B routine, described in the discussion of Case 1, would then be required.

Note that if the original design calculations were done on the assumption that reflux is not subcooled, the effect of subcooling may be considered as a change in L_R :

$$\Delta L_R = \frac{c_p}{\lambda}(T_o - T_R)$$

If the base case assumes some subcooling ($T_R = T_{R1}$), and later the subcooling is different ($T_R = T_{R2}$):

$$\Delta L_R = \frac{c_p}{\lambda}(T_{R2} - T_{R1})$$

Case 4

For the control system of Figure 19.3, if feed and reflux are fixed, and if it is desired to find the responses x'_D, x'_B to changes in boilup, the following "prep" equations are required:

$$B' = B - \Delta V_s$$

$$V'_s = V_s + \Delta V_s$$

$$\beta' = V'_s/B'$$

$$D' = D + \Delta V_s$$

$$R' = L_R/D'$$

The Type B routine, described in discussion of Case 1, is then required.

Case 5

A third basic control scheme is that of Figure 19.4. Here overhead condensate receiver level is controlled by throttling distillate flow, while reflux is flow controlled, perhaps ultimately cascaded from overhead composition control. Base level is controlled by adjusting heating-medium flow control. Bottom-product flow is on flow control, perhaps ultimately cascaded from bottom-product composition control.

If feed and bottom rates are fixed, we wish to find the responses x'_D, x'_B to a change in reflux flow. The following "prep" equations are required:

$$L'_R = L_R + \Delta L_R$$

$$V'_s = V_s + \Delta L_R$$

$$\beta' = V'_s/B$$

$$R' = L'_R/D$$

The Type B routine, described in the discussion of Case 1, is required.

Case 6

If, for Figure 19.4, feed and reflux are fixed, we may wish to find the responses x'_D, x'_B to a change in bottom-product flow. The following “prep” equations are required:

$$B' = B + \Delta B$$

$$V'_s = V_s - \Delta B$$

$$\beta' = V'_s/B'$$

$$D' = D - \Delta B$$

$$R' = L_R/D'$$

The Type B routine, described under Case 1, is required.

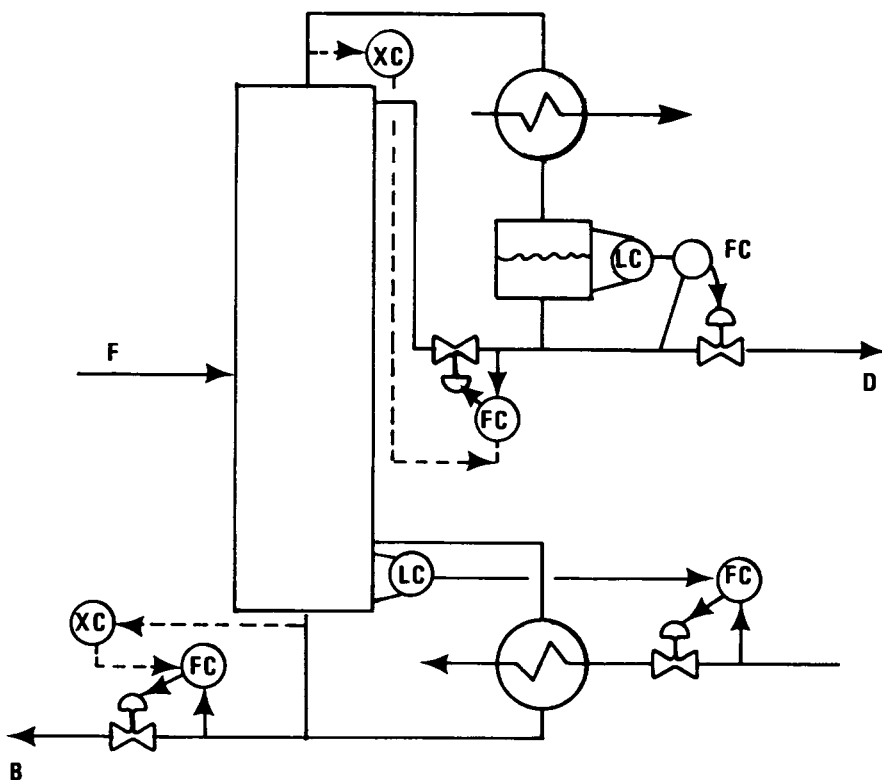


FIGURE 19.4

Distillate via reflux drum level control; boil up via base level control

Column Terminal Composition Sensitivity to Various Inputs**Case 7**

For the control system of Figure 19.2, we may wish to determine overhead and base composition responses to changes in feed rate. The following “prep” equations are needed:

$$B' = \frac{B + \Delta F}{1 + \Delta F}$$

$$\beta' = V_s/B'$$

$$L'_R = L_R + (1 - q)\Delta F$$

$$D' = 1 - B'$$

$$R' = L'_R/D'$$

In the first equation above, B' might better be labeled B'' . Since D is fixed:

$$B' = B + \Delta F$$

But our program requires that all flows be relative to $F = 1$. Therefore let:

$$B' = \frac{B + \Delta F}{F + \Delta F} = \frac{B + \Delta F}{1 + \Delta F}$$

For simplicity of symbolism, we represent B'' by B' .

Case 8

For the control system of Figure 19.2, we may wish to find the responses x'_D and x'_B to changes in q . The following “prep” equations are needed:

$$q' = q + \Delta q$$

$$L'_R = L_R - (\Delta q)F$$

$$R' = L'_R/D$$

Use the Type B program as described under Case 1.

Case 9

For the control system of Figure 19.3, we may wish to determine overhead and base composition responses to changes in feed rate. The following “prep” equations are needed:

$$B' = \frac{B + q\Delta F}{1 + \Delta F}$$

$$\beta' = V_s/B'$$

$$D' = 1 - B'$$

$$R' = L_R/D'$$

In the first equation above, B' again might more properly be written B'' . For a change in F to $F' = F + \Delta F$:

$$B' = L_R + q(F + \Delta F) - V_s$$

But:

$$B = L_R + qF - V_s$$

and therefore:

$$B' - B = q\Delta F$$

or

$$B' = q\Delta F + B$$

Since our programs are based on 1 mole feed per unit time, let us define:

$$B'' = \frac{q\Delta F + B}{F + \Delta F}$$

$$B'' = \frac{q\Delta F + B}{1 + \Delta F}$$

Again, for simplicity of symbolism, we use B' instead of B'' .

A Type B program, discussed under Case 1, should be used.

Case 10

For the control system of Figure 19.3, we may wish to find the responses of x_D and x_B to changes in q . The following “prep” equations are required:

$$q' = q + \Delta q$$

$$B' = B + (\Delta q)F$$

$$\beta' = V_s/B'$$

$$D' = D - (\Delta q)F$$

$$R' = L_R/D'$$

The Type B program (see Case 1) should then be employed.

Case 11

For the control system of Figure 19.4, we may wish to determine overhead and base composition responses to changes in feed rate. The following “prep” equations are required:

$$B' = \frac{B}{1 + \Delta F}$$

$$V'_s = V_s + q\Delta F$$

$$\beta' = V'_s/B'$$

$$D' = 1 - B'$$

$$R' = L_R/D'$$

Use the Type B program (see Case 1).

Case 12

For the control system of Figure 19.4, we may wish to determine overhead and base composition responses to changes in feed enthalpy factor, q . The following “prep” equations are needed:

$$q' = q + \Delta q \quad F = 1.00$$

$$V'_s = V_s + (\Delta q)F$$

$$\beta' = V'_s/B$$

Use the Type B program (see Case 1).

Case 13

For any of the control schemes, the effect of feed composition change from z_F to z'_F on x_D and x_B may be found simply by entering the new z'_F into data storage for the Type B program (see Case 1).

Case 14

For any of the control schemes, the effect of a column pressure change from P to P' on x_D and x_B may be found by recalculating α 's; rerunning the program for the quadratic coefficients A_α , B_α , and C_α ; and entering the new values of A_α , B_α , and C_α in the Type B program (see Case 1).

Column Gains for Feedforward Compensation

Here the objective is to find required feedforward compensator gain to hold terminal composition constant as various external factors vary.

Case 15

For any of the control systems discussed—Figure 19.2, 19.3, or 19.4—assume that top composition and bottom composition are held constant. We wish to find the changes in L_R , D , B , and V_s required to hold compositions constant in the face of a feed composition change, Δz_F . This information could be used to design feedforward compensators to minimize transient changes in terminal compositions. The variables chosen for feedforward compensation will depend on which feedback control scheme is used—Figure 19.2, 19.3, or 19.4.

The equations required to recalculate the other variables are:

$$z'_F = z_F + \Delta z_F \quad F = 1.00$$

$$D' = D + \frac{F \Delta z_F}{x_D - x_B}$$

$$B = F - D'$$

Use the Type A program, described in Section 19.3.

Case 16

This is the same as Case 15 except that q is changed. The only equation needed is:

$$q' = q + \Delta q$$

Use the Type A program (see Section 19.3).

Case 17

This is the same as Case 15 except that column top pressure is changed to $P' = P + \Delta P$. Relative volatilities should be recalculated and the quadratic coefficients A_α , B_α , and C_α reevaluated. Then run the Type A program (see Section 19.3).

Economic Penalty of Overrefluxing

Case 18

If product purity at one end of the column, say x_D , is set at a new, constant, higher value, to ensure that product purity is always at least as good as the original specification, we may wish to estimate the cost penalty of the increase boilup. We will need to know L_R and V_s . If the column control scheme fixes either D or B , the other will also be fixed. Then if one new terminal composition is chosen, the other is readily calculated:

$$x'_D = x_D + \Delta x_D$$

$$x'_B = x_B - x_D \left(\frac{D}{B} \right)$$

Use the Type A program (see Section 19.3).

Case 19

If we have a situation similar to that of Case 18 except that both a new x'_D and a new x'_B are chosen, we will get changes in D and B as well as L_R and V_s :

$$x'_D = x_D + \Delta x_D \quad B'/F' = \frac{z_F - x'_D}{x'_B - x'_D}$$

$$x'_B = x_B + \Delta x_B \quad D' = F - B'$$

Use the Type A program (see Section 19.3).

Shinskey⁷ has pointed out that the differences between the design value of V_s/F and those required for higher values of x_D are measures of the cost of overrefluxing.

Case 20

If a column has automatic control of $R = L_R/D$ (it does not matter whether condensate receiver is level controlled via reflux or top product), the following “prep” equations apply for constant V_s :

$$R' = R + \Delta R$$

$$L'_R = R'D'$$

$$L'_R + D' = L_R + D$$

or

$$R'D' + D' = RD + D$$

so:

$$D' = \left(\frac{1 + R}{1 + R'} \right) D$$

$$B' = 1 - D'$$

$$\beta = V_s/B'$$

Use the Type B procedure (see Case 1).

Case 21

For a column with automatic control of $R = L/D$, the following “prep” equations apply for constant R :

$$V'_s = V_s + \Delta V_s$$

$$D' + L'_R = V'_R = V'_s (1 - q)$$

But:

$$L'_R = RD'$$

so:

$$D' + RD' = V'_s (1 - q)$$

or

$$D' = \frac{V'_s (1 - q)}{1 + R}$$

$$1 - D' = B'$$

$$V'_s/B' = \beta'$$

Use the Type B procedure (see Case 1).

19.5 EXAMPLES

Example 1: Calculation of Column Gains

Let us use a test case where the following conditions apply:

$$z_F = 0.46$$

$$x_D = 0.98$$

$$x_B = 0.01$$

$$q = 0.7$$

$$\alpha = 2.80 \text{ at } x = 0.20$$

$$\alpha = 2.20 \text{ at } x = 0.80$$

$$R = 2.50$$

Our HP-97 program predicts:

$$N_s = 7$$

$$N_T = 16$$

$$(N_R = 9)$$

$$\beta = 2.469$$

$$(x_D)_{\text{final}} = 0.9930$$

Next we found the exact R to be 2.306. We then chose to find the column gains, dx_D/dD and dx_B/dD , which is Case 1.

After running the design and exact R (Type A) cases, we made a data card. With this card, we ran a Type B program twice, once for D slightly smaller than design, and once for D slightly larger than design. V_s was held constant. We fitted the two sets of three points—one set for x_D and one set for x_B —to a quadratic function:

$$y = Ax^2 + Bx + C$$

$$\frac{dy}{dx} = 2A\bar{x} + B$$

The results are tabulated in Table 19.1. We arbitrarily chose $F = 1.00$, so for the design case, $D = 0.46392$ and $B = 0.53608$. Although we chose small changes, $\Delta D = 0.001$, we found large changes in gain with changes in D . The gain is a very nonlinear function, decreasing as x_D increases and as D decreases. It is also a function of F . One could use this information to design a suitable nonlinear or adaptive control scheme.

Example 2: Economic Penalty of Overrefluxing

Douglas and Seemann¹ have investigated the consequences of overrefluxing a deisobutanizer. Design conditions were:

$$z_F = 0.575$$

$$x_D = 0.950$$

$$x_B = 0.050$$

$$q = 1.00$$

$$\alpha = 1.332$$

$$R = 5.87$$

They assumed a control system in which distillate is flow controlled, reflux is set by condensate receiver (reflux drum) level control, and bottom-product rate is controlled by column-base level. Overhead composition is controlled by boilup.

For the design case, Douglas and Seemann calculated by their analytical method, based on a simplified Smoker equation, that $N_i = 26.30$ stages, including the reboiler, and $N_R = 31.65$ trays. The exact Smoker equation, as well as the tray-to-tray method, predicts $N_i = 21$ stages and $N_R = 18$ stages. Predicted overhead composition by our HP-97 program is 0.9520. The exact R turns out to be 5.852. Again we made a data card at this point to minimize subsequent calculations.

Table 19.2 shows the results obtained with various values of x_D greater than the design value of 0.950.

TABLE 19.1

Variation of terminal compositions and gains with distillate rate for a constant boilup rate (gains on basis 1.00 mols feed)

D	x_D	x_B	$\frac{dx_D}{dD}$	$\frac{dx_B}{dD}$
0.46292	0.98113	0.010835	-1.063	-0.891
0.46392	0.98000	0.01000	-1.189	-0.780
0.46492	0.97875	0.009273	-1.314	-0.670

TABLE 19.2

Energy increase due to over-refluxing deisobutanizer

TOP PRODUCT COMPOSITION x_D	0.950	0.955	0.960	0.965	0.970
BOTTOM PRODUCT COMPOSITION x_B	0.050	0.043	0.036	0.029	0.022
REFLUX RATIO $R = L_R/D$ TRAY-TO-TRAY	5.8521	6.1880	6.6145	7.1834	8.0185
VAPOR BOILUP-TO-FEED RATIO V_S/F TRAY-TO-TRAY	3.997	4.193	4.442	4.774	5.261
% ENERGY INCREASE TRAY-TO-TRAY	0	4.93	11.13	19.44	31.62
REFLUX RATIO $R = L_R/D$ DOUGLAS-SEEMANN ANALYTICAL METHOD	5.87	5.98	6.12	6.33	6.65
% ENERGY INCREASE DOUGLAS-SEEMANN ANALYTICAL METHOD	0	1.61	3.70	6.72	11.47

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20 Composition Control—Binary Distillation

20.1 INTRODUCTION

a truly definitive treatment of composition control, even for simple binary distillation, has not yet been published—and the reader will not find one here. Some very interesting papers have been published that demonstrate exotic techniques, usually involving “modern control theory.” These papers are largely aimed at multivariable control, a laudable objective. Unfortunately most of the studies are concerned with product impurity levels that are typically in the range of 1–5 percent and the authors assume constant relative volatility and linear column models. Real-life industrial columns are often highly nonideal (nonlinear) and often must produce products with high purities—0.1 mol percent impurities or less. The last consideration results in very nonlinear behavior.¹ For simplicity’s sake, however, we will assume linearity in this chapter to give some insight into column-composition control principles. But particularly for composition control at both ends of the column, we strongly recommend a simulation study using a nonlinear model to compare different control schemes.

Some very perceptive insights into the differences between petroleum refining distillation and that in the chemical industry are offered by Tolliver and Waggoner² in an extensive literature review.

Although most existing columns do not have composition control at both ends, some columns are so equipped. Most of those reported in the literature (references 8 and 9, for example) apparently are not badly afflicted with interaction. At least dual composition control was accomplished without decouplers. For a laboratory column Waller and associates¹⁰ studied six approaches to dual composition control, including those of modern control theory. Tyreus¹¹ discusses multivariable control of an industrial column whose design was performed via a technique called the inverse Nyquist array.

20.2 FEEDBACK CONTROL OF COMPOSITION

In Chapter 18 overall composition dynamics, based on the Rippin and Lamb model, were represented graphically in Figure 18.1. If we assume that boilup, bottom-product flow, reflux, and top-product flow are the manipulated flows, then for the system of Figure 20.1 we may prepare the closed-loop signal flow diagram of Figure 20.2 for:

1. Condensate receiver level control cascaded to distillate flow control
2. Overhead composition control cascaded to reflux flow control
3. Base level control cascaded to bottom-product flow control
4. Base composition control cascaded to steam flow control

For all four loops we assume that the secondary flow controls are fast compared with the primary level and composition loops. This permits us to ignore flow control loop dynamics. Then to go from the primary controller output to flow we need only multiply by $1/K_{mf}$ where K_{mf} is the flow-meter static gain (linear flow meters assumed).

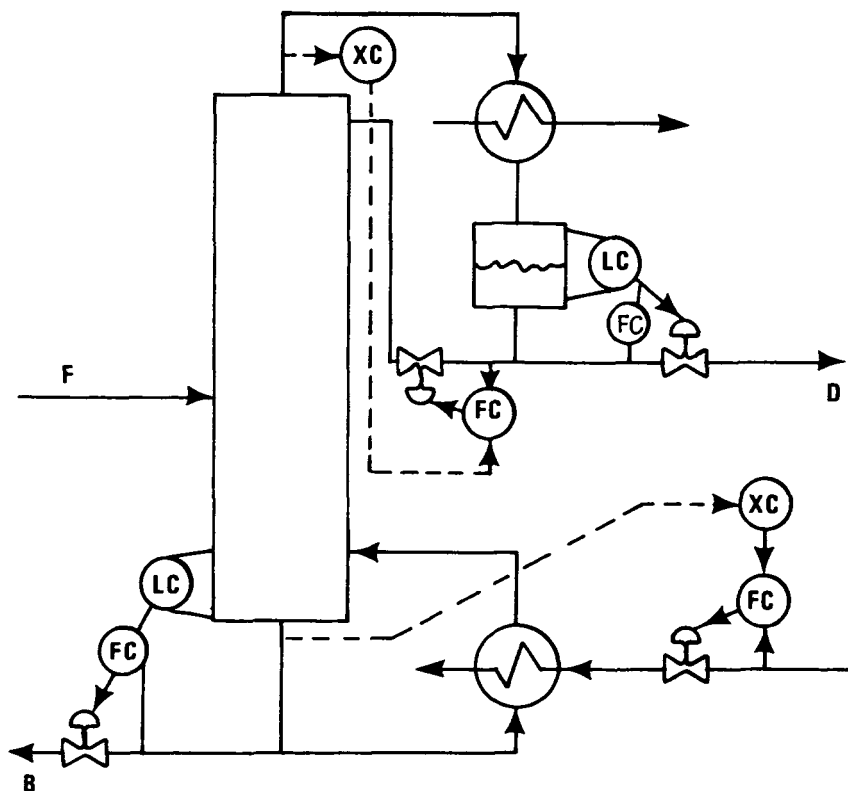


FIGURE 20.1

Distillate via reflux drum level control bottom product via base level control

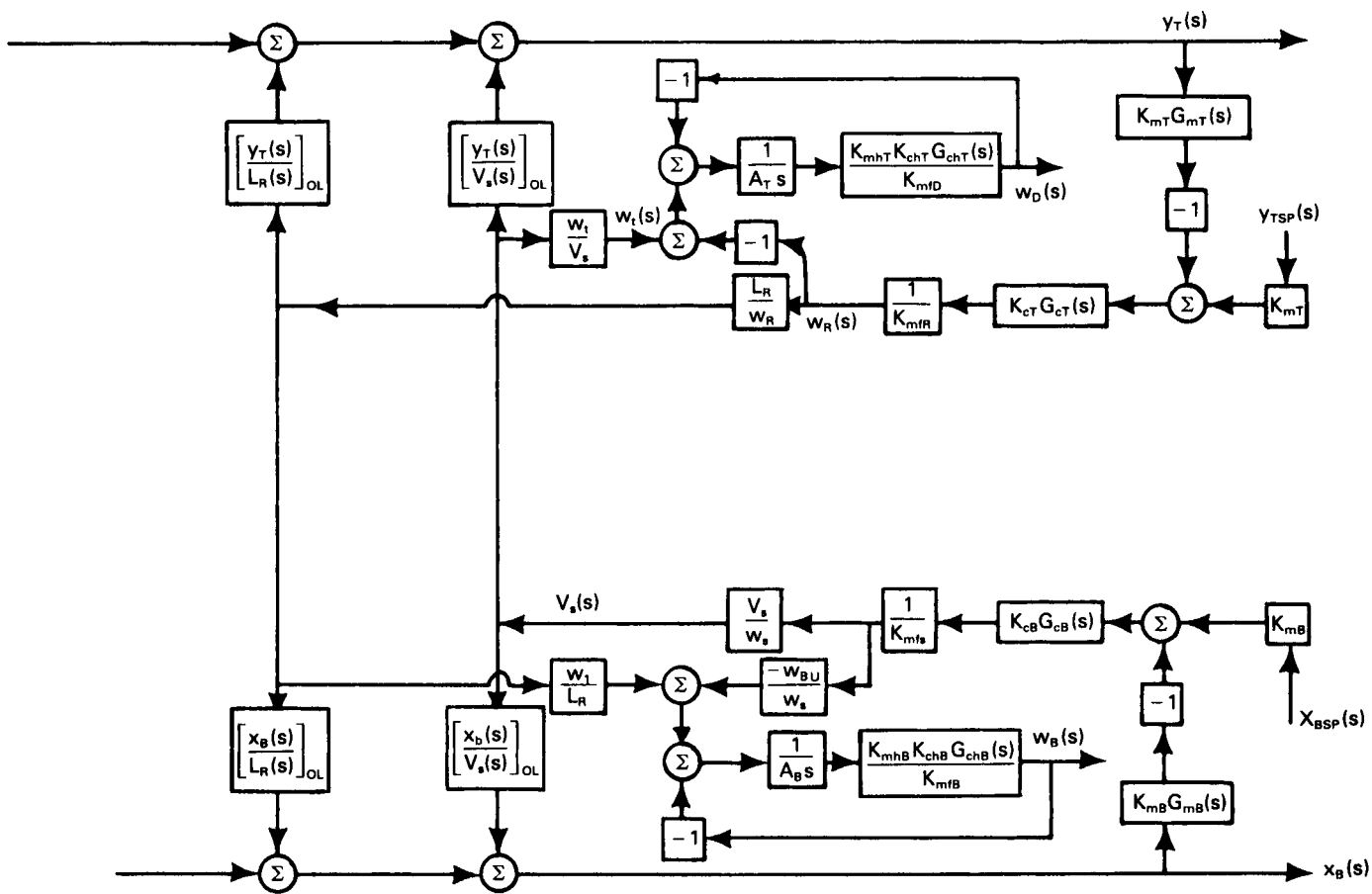


FIGURE 20.2
Partial signal flow diagram for figure 20.1

Figure 20.2 shows that the two level control loops are not nested in the composition control loops. If, however, we cascade (1) top composition control to distillate flow control and (2) condensate receiver level control to reflux flow control (Shinskey's "material-balance control"), then, as shown by the signal flow diagram of Figure 20.3, the top level control is "nested" in the top composition control loop. Since column composition does not change until reflux flow changes, we must use tight level control. This means that the condensate receiver may not be used for distillate flow smoothing to the next process step. It also means that if the operator puts the receiver level control loop on "manual," there is no composition control. Since none of our studies have ever turned up a case where this scheme offered advantages for composition controls, we use it only occasionally for very high reflux ratio columns.

20.3 INTERACTION COMPENSATION

In Chapter 12 we proposed a particular control-loop structure that incorporates overrides and antireset windup. It also accommodates feedforward compensation and advanced control techniques without interfering with either normal reset or antireset windup. We also suggested that decouplers could be designed to compensate for interactions in the same way that feedforward compensators are designed. The technique here leads to stable, noninteracting control, but not necessarily to optimum control. Modern control theory, with its more sophisticated approaches to multivariable control, sometimes requires some interaction for optimality.

A possible implementation is shown in Figure 20.4 for the system of Figure 20.2. The decoupler for canceling the effect of top composition controller output changes on the bottom composition has the transfer function:

$$\theta_{FT}(s) = K_{FT}G_{FT}(s)\left(\frac{\tau_{RB}s}{\tau_{RB}s + 1}\right)\theta_{CT}(s) \quad (20.1)$$

Note that the feedback controller, $K_{CT}G_{CT}(s)$, is shown as a PI controller with external reset feedback.

The decoupler for canceling the effect of bottom composition controller output changes on top composition has the transfer function:

$$\theta_{FB}(s) = K_{FB}G_{FB}(s)\left(\frac{\tau_{RT}s}{\tau_{RT}s + 1}\right)\theta_{CB}(s) \quad (20.2)$$

Note that the impulse function time constants are the same as the reset time constants of the loops to which the decouplers are connected.

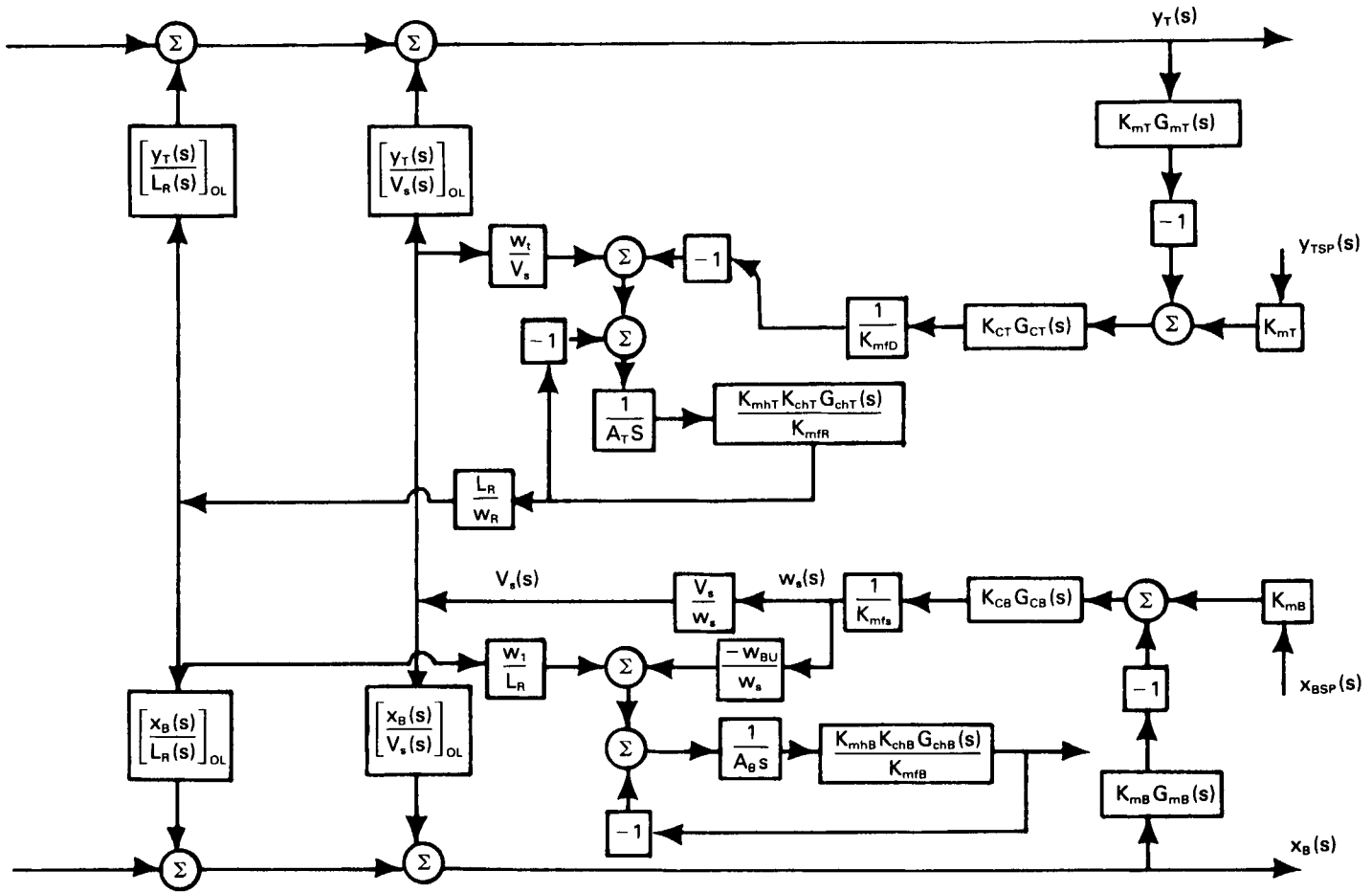


FIGURE 20.3
Partial signal flow diagram for system with reflux manipulated by reflux drum level

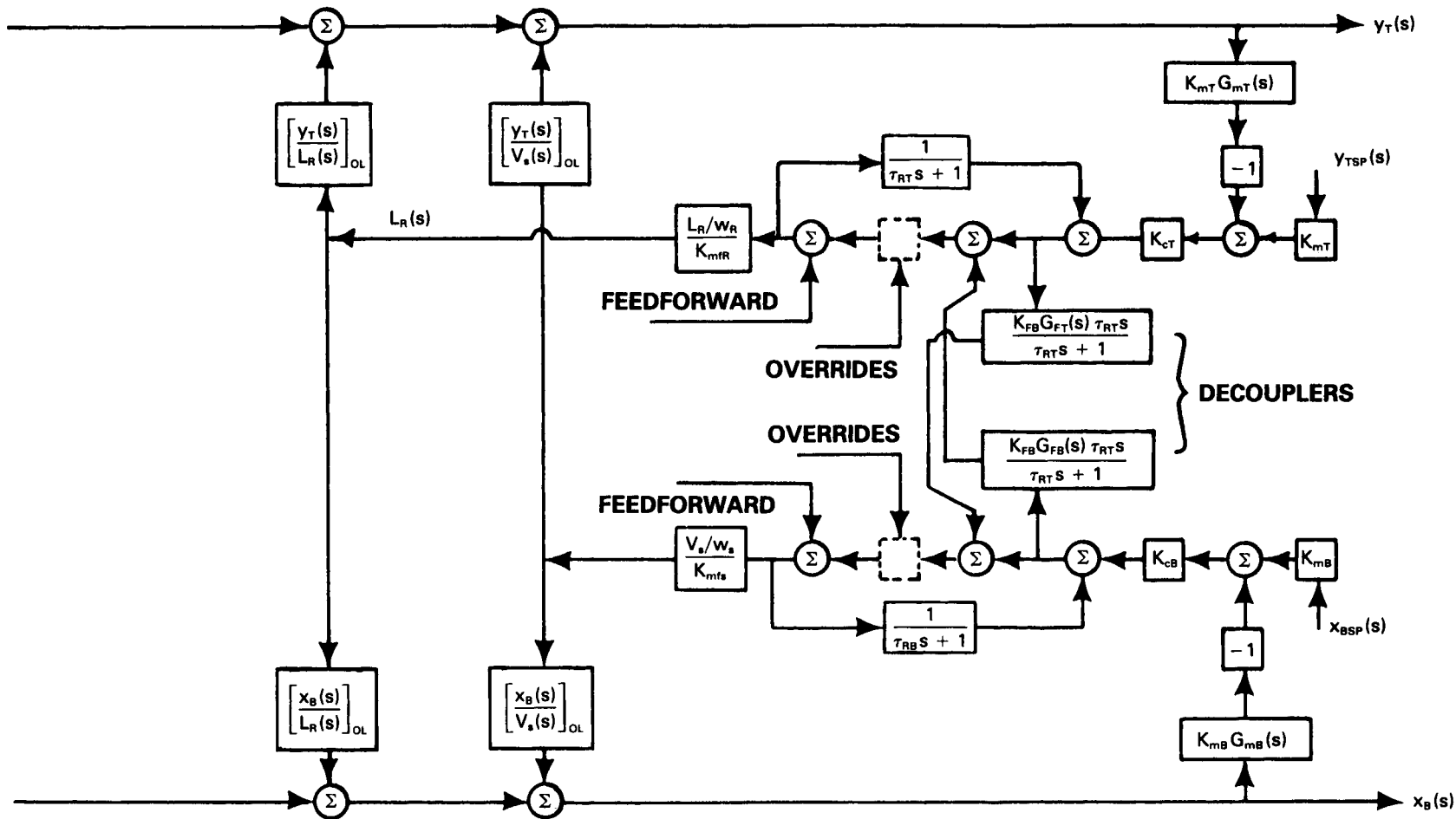


FIGURE 20.4
Signal flow diagram for system of figure 20.2 with decouplers

The signal flow diagram of Figure 20.4 may now be reduced by means of signal flow diagram transformation theorems to the form of Figure 20.5. By inspection we can see that the effect of reflux flow changes on bottom composition, x_B , may be canceled out if we make:

$$\frac{L_R/w_R}{K_{mfR}} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} + K_{FT} G_{FT}(s) \frac{V_s/w_s}{K_{mfs}} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} = 0 \quad (20.3)$$

while the effect of vapor flow changes on top composition, x_T , may be canceled out if we make:

$$\frac{V_s/w_s}{K_{mfs}} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} + K_{FB} G_{FB}(s) \frac{L_R/w_R}{K_{mfR}} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} = 0 \quad (20.4)$$

In solving equation (20.3) we find that the compensator, $K_{FT} G_{FT}(s)$, has the transfer function:

$$K_{FT} G_{FT}(s) = \frac{-\frac{L_R/w_R}{K_{mfR}} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL}}{\frac{V_s/w_s}{K_{mfs}} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL}} \quad (20.5)$$

We can also find the other compensator transfer function from equation (20.4):

$$K_{FB} G_{FB}(s) = \frac{-\frac{V_s/w_s}{K_{mfs}} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL}}{\frac{L_R/w_R}{K_{mfR}} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL}} \quad (20.6)$$

Since the two loops are now decoupled, we may prepare the partial signal flow diagram of Figure 20.6 that shows that the two loops are now independent. By substituting equation (20.5) into the large block of the top loop, we get:

$$\begin{aligned} & \frac{L_R/w_R}{K_{mfR}} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} + \left(\frac{-\frac{L_R/w_R}{K_{mfR}} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL}}{\frac{V_s/w_s}{K_{mfs}} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL}} \right) \times \frac{V_s/w_s}{K_{mfs}} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \\ &= \frac{L_R/w_R}{K_{mfR}} \left(\left[\frac{y_T(s)}{L_R(s)} \right]_{OL} - \frac{\left[\frac{x_B(s)}{L_R(s)} \right]_{OL}}{\left[\frac{x_B(s)}{V_s(s)} \right]_{OL}} \times \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right) \end{aligned} \quad (20.7)$$

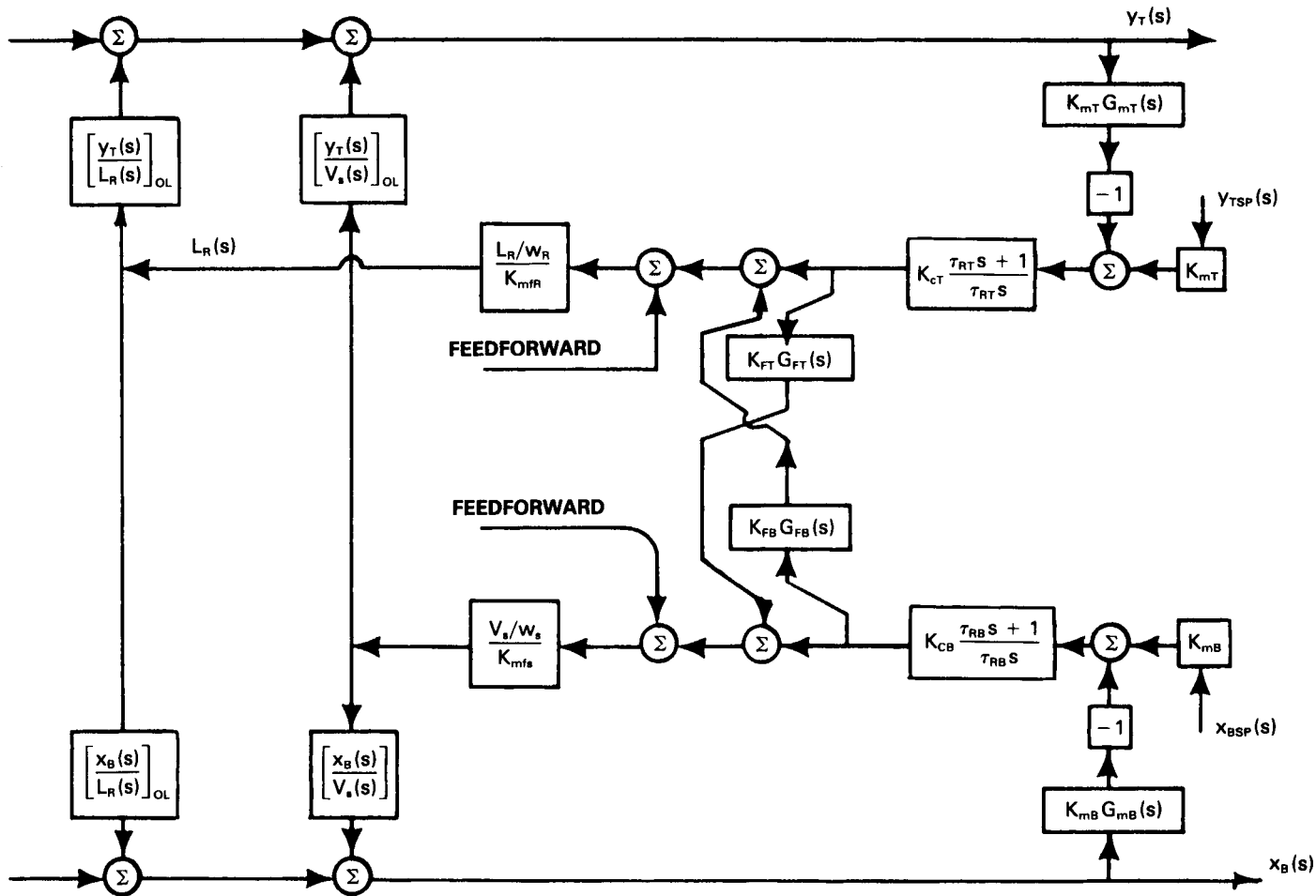


FIGURE 20.5
Partly reduced signal flow diagram of figure 20.4

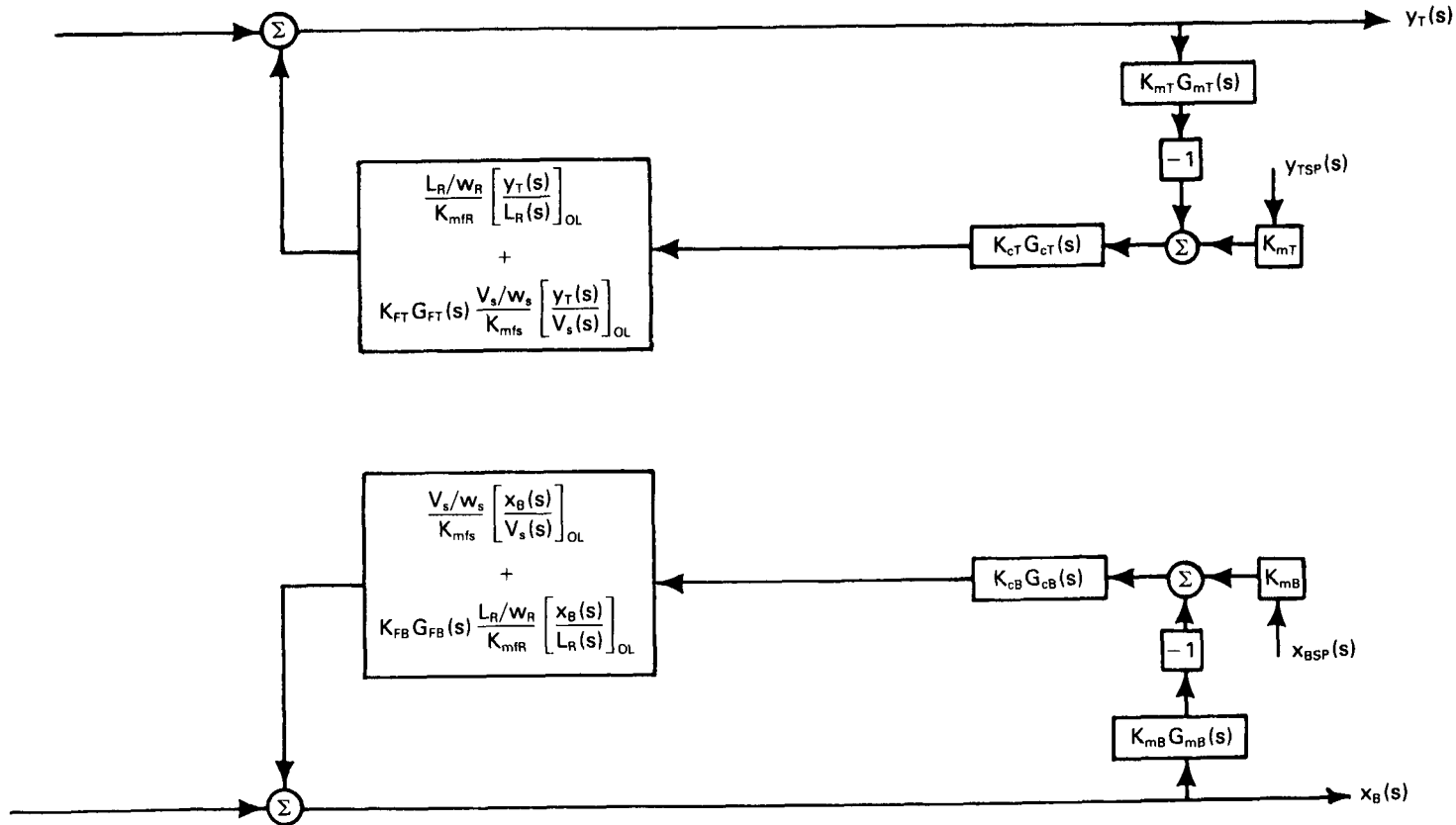


FIGURE 20.6
Partly reduced signal flow diagram of figure 20.5

Similarly, by substituting equation (20.6) into the bottom loop, we obtain:

$$\begin{aligned} \frac{V_s/w_s}{K_{mfS}} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} + \left(\frac{-\frac{V_s/w_s}{K_{mfS}} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL}}{\frac{L_R/w_R}{K_{mfR}} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL}} \right) \times \frac{L_R/w_R}{K_{mfR}} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \\ = \frac{V_s/w_s}{K_{mfS}} \left(\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} - \frac{\left[\frac{y_T(s)}{V_s(s)} \right]_{OL}}{\left[\frac{y_T(s)}{L_R(s)} \right]_{OL}} \times \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \right) \quad (20.8) \end{aligned}$$

As an example let us consider a binary distillation column designed to separate water from nitric acid. Top composition is controlled by manipulating reflux (see Figure 20.1) while the base composition is controlled by adjusting steam flow. The following transfer functions were derived via the stepping technique:

$$\left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \simeq \frac{0.000207 e^{-3.6s}}{31.5 s + 1} \quad (20.9)$$

$$\left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \simeq \frac{-0.0000036}{21.6 s + 1} \quad (20.10)$$

$$\left[\frac{y_T(s)}{L_R(s)} \right]_{OL} \simeq \frac{0.0000097}{21.6 s + 1} \quad (20.11)$$

$$\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \simeq \frac{-0.000215}{28 s + 1} \quad (20.12)$$

The time constants are in minutes.

From these we find that the large top block of Figure 20.6 becomes [equation (20.7)]:

$$\begin{aligned} \left(\frac{L_R/w_R}{K_{mfR}} \right) \left(\left[\frac{y_T(s)}{L_R(s)} \right]_{OL} - \frac{\left[\frac{x_B(s)}{L_R(s)} \right]_{OL}}{\left[\frac{x_B(s)}{V_s(s)} \right]_{OL}} \times \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right) \\ \simeq \left(\frac{L_R/w_R}{K_{mfR}} \right) \left(\frac{0.0000062}{21.6 s + 1} \right) \quad (20.13) \end{aligned}$$

Correspondingly, the large block in the bottom loop of Figure 20.6 [Equation (20.8)] becomes:

$$\left(\frac{V_s/w_s}{K_{mfs}}\right) \left(\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} - \frac{\left[\frac{y_T(s)}{V_s(s)} \right]_{OL}}{\left[\frac{y_T(s)}{L_R(s)} \right]_{OL}} \times \left[\frac{x_B(s)}{L_R(s)} \right] \right) = \left(\frac{V_s/w_s}{K_{mfs}}\right) \left(\frac{0.00018}{28.1s + 1} \right) \quad (20.14)$$

The decouplers themselves become:

$$K_{FT}G_{FT}(s) \cong 0.36 e^{-3.6s} \cong 0.36 \left(\frac{1 - 1.8s}{1 + 1.8s} \right) \quad (20.15)$$

$$K_{FB}G_{FB}(s) \cong 1.02 \quad (20.16)$$

Equations (20.13) and (20.14) indicate simpler process dynamics than probably exists in reality. But we would expect, in any event, that a PI controller with fairly high gain could be used. The decouplers for this problem are easily implemented with analog pneumatic or electronic devices.

Since the column as built had a composition analyzer at only one end, we had no opportunity to test operation with decouplers.

Luyben¹² has discussed two approaches to decoupling, including one he terms "ideal," but that offers limited benefits and is more complex to implement. Later Luyben and Vinante¹³ tested decoupling experimentally on a semiworks column. Since the column had little interaction, benefits of decoupling were minimal. Niederlinski¹⁴ and Waller¹⁵ have also studied decoupling.

20.4 FEEDFORWARD COMPENSATION

Feedforward compensators may be added for almost any disturbance, as shown by Figure 20.7. Physically each may be implemented, as discussed in Chapter 12, with an impulse function and a summer connected inside of the composition feedback controller reset circuit.

Individual feedforward compensator functions may be determined in a very simple fashion. Let us, for example, look at feedforward compensation for feed-rate changes, F . If the compensators do a perfect job, then $y_T(s) = 0$ and $x_B(s) = 0$, and there will be no contribution from the feedback controllers. Then, to make $y_T(s) = 0$, we can see by inspection of Figure 20.7 that:

$$F(s) \left(\left[\frac{y_T(s)}{F(s)} \right]_{OL} + K_{f4}G_{f4}(s) \times \alpha \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} + K_{f2}G_{f2}(s) \times \beta \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right) = 0 \quad (20.17)$$

where

$$\alpha = \frac{L_R/w_R}{K_{mfR}} \quad (20.18)$$

and

$$\beta = \frac{V_s/w_s}{K_{mfs}} \quad (20.19)$$

Similarly, to make $x_B(s) = 0$, we may write:

$$F(s) \left(\left[\frac{x_B(s)}{F(s)} \right]_{OL} + K_{f2} G_{f2}(s) \times \beta \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} + K_{f4} G_{f4}(s) \times \alpha \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \right) = 0 \quad (20.20)$$

By simultaneous solution of equations (20.17) and (20.18) we find that

$$K_{f2} G_{f2}(s) = \frac{\left[\frac{y_T(s)}{F(s)} \right]_{OL} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} - \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} \left[\frac{x_B(s)}{F(s)} \right]_{OL}}{\beta \left(\left[\frac{y_T(s)}{L_R(s)} \right]_{OL} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} - \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right)} \quad (20.21)$$

and

$$K_{f4} G_{f4}(s) = \frac{\left[\frac{x_B(s)}{F(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} - \left[\frac{y_T(s)}{F(s)} \right]_{OL} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL}}{\alpha \left(\left[\frac{y_T(s)}{L_R(s)} \right]_{OL} \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} - \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right)} \quad (20.22)$$

By a similar analysis we find that, to make $y_T(s) = 0$ and $x_B(s) = 0$ in response to disturbances $z_F(s)$ and $q(s)$, we can derive the following feedforward compensator functions:

$$K_{f1} G_{f1}(s) = \frac{\left[\frac{y_T(s)}{z_F(s)} \right]_{OL} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} - \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} \left[\frac{x_B(s)}{z_F(s)} \right]_{OL}}{\beta \left(\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} - \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right)} \quad (20.23)$$

$$K_{f3} G_{f3}(s) = \frac{\left[\frac{x_B(s)}{z_F(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} - \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \left[\frac{y_T(s)}{z_F(s)} \right]_{OL}}{\alpha \left(\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} - \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right)} \quad (20.24)$$

$$K_{f5}G_{f5}(s) = \frac{\left[\frac{y_T(s)}{q(s)} \right]_{OL} \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} - \left[\frac{x_B(s)}{q(s)} \right]_{OL} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL}}{\beta \left(\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} - \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right)} \quad (20.25)$$

$$K_{f6}G_{f6}(s) = \frac{\left[\frac{x_B(s)}{q(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} - \left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \left[\frac{y_T(s)}{q(s)} \right]_{OL}}{\alpha \left(\left[\frac{x_B(s)}{V_s(s)} \right]_{OL} \left[\frac{y_T(s)}{L_R(s)} \right]_{OL} - \left[\frac{x_B(s)}{L_R(s)} \right]_{OL} \left[\frac{y_T(s)}{V_s(s)} \right]_{OL} \right)} \quad (20.26)$$

In practice, since feed flow rate changes constitute by far the major disturbances to most distillation columns, there has been a tendency to concentrate on compensating for them. Further, it has been found that major improvement results from using only static feedforward; the incremental improvement obtained with the various $G(s)$ terms is comparatively small. Usually, therefore, one finds in practice a static gain term with a first-order lag or simple lead-lag dynamic compensation. It is likely, however, that more exact dynamic functions will be beneficial in at least some cases.

Some of the practical problems in feeding forward from feed compensation have been discussed by Luyben.³ Luyben⁴ has also discussed the problems caused by inverse response in the design of feedforward compensators as well as feedback controllers for composition.

For columns that are fairly nonideal, some sort of on-line identification procedure is necessary to tune the various feedforward parameters adaptively.

20.5 RELATIVE-GAIN MATRIX

The relative-gain matrix, or relative-gain array (RGA), was originally suggested by Bristol¹⁹ as a means of determining the steady-state interaction between process variables. Shinskey⁵ and McAvoy⁶ have been assiduous in exploring the applications to distillation columns. One of the most lucid and concise treatments we have seen is that of Ray.⁷ The implication is that by proper "pairing" of variables one may arrive, in some instances, at a control loop structure that promises less interaction than other feasible structures.

Like other "linear" techniques, the relative-gain array assumes that the principle of superposition holds. In the published papers and books there is also an assumption, not implicit in the mathematics, that the steady-state gains give a true indication of interaction. For real systems dynamics effects may be just the opposite of steady-state effects and may be dominant. Further, since a distillation column is apt to be just one equipment piece in a sequence of process steps, there are usually fewer choices of control system structure than

implied by the literature. Finally, if one is going to use decouplers, as is entirely practical today, there may be little preference between various control system structures.

Some work has been done on defining a dynamic RGA. See reference 6.

Relative-Gain Matrix for Binary Column—I

Let us consider our normally preferred scheme:

1. Condensate receiver level controlled by distillate.
2. Top composition controlled by reflux.
3. Bottom composition controlled by boilup (usually steam flow).
4. Base level controlled by bottom product withdrawal.

By the principle of superposition we may write two equations:

$$\Delta y_T = \frac{\partial y_T}{\partial L_R} \Delta L_R + \frac{\partial y_T}{\partial V_s} \Delta V_s \quad (20.27)$$

and

$$\Delta x_B = \frac{\partial x_B}{\partial L_R} \Delta L_R + \frac{\partial x_B}{\partial V_s} \Delta V_s \quad (20.28)$$

whence, in matrix form:

$$\begin{bmatrix} y_T \\ x_B \end{bmatrix} = \begin{bmatrix} \frac{\partial y_T}{\partial L_R} & \frac{\partial y_T}{\partial V_s} \\ \frac{\partial x_B}{\partial L_R} & \frac{\partial x_B}{\partial V_s} \end{bmatrix} \begin{bmatrix} L_R \\ V_s \end{bmatrix}$$

The relative-gain array is related to the first term in brackets on the right-hand side of equation (20.29):

$$\Lambda = \begin{bmatrix} \lambda_1 & \lambda_2 \\ \lambda_3 & \lambda_4 \end{bmatrix} \quad (20.29a)$$

According to the theory:

$$\lambda_1 = \frac{\left. \frac{\Delta y_T}{\Delta L_R} \right|_{\text{constant } V_s}}{\left. \frac{\Delta y_T}{\Delta L_R} \right|_{\text{constant } x_B}}$$

If V_s is constant, $\Delta V_s = 0$ and from equation (20.27) we get:

$$\left. \frac{\Delta y_T}{\Delta L_R} \right|_{\text{constant } V_s} = \frac{\partial y_T}{\partial L_R} \quad (20.30)$$

If x_B is constant, then $\Delta x_B = 0$ and from equation (20.28) we get:

$$\Delta V_s = \frac{-\frac{\partial x_B}{\partial L_R} \Delta L_R}{\frac{\partial x_B}{\partial V_s}} \quad (20.31)$$

On substituting this back into equation (20.27), we obtain:

$$\Delta y_T = \frac{\partial y_T}{\partial L_R} \Delta L_R + \frac{\partial y_T}{\partial V_s} \left[\frac{-\frac{\partial x_B}{\partial L_R} \Delta L_R}{\frac{\partial x_B}{\partial V_s}} \right] \quad (20.32)$$

or

$$\left. \frac{\Delta y_T}{\Delta L_R} \right|_{\text{constant } x_B} = \frac{\partial y_T}{\partial L_R} + \frac{\partial y_T}{\partial V_s} \left[\frac{-\frac{\partial x_B}{\partial L_R}}{\frac{\partial x_B}{\partial V_s}} \right] \quad (20.33)$$

Hence:

$$\lambda_1 = \frac{\frac{\partial y_T}{\partial L_R}}{\frac{\partial y_T}{\partial L_R} + \frac{\partial y_T}{\partial V_s} \left[\frac{-\frac{\partial x_B}{\partial L_R}}{\frac{\partial x_B}{\partial V_s}} \right]} \quad (20.34)$$

Next:

$$\lambda_2 = \frac{\left. \frac{\Delta y_T}{\Delta V_s} \right|_{\text{constant } L_R}}{\left. \frac{\Delta y_T}{\Delta V_s} \right|_{\text{constant } x_B}} = \frac{\frac{\partial y_T}{\partial V_s}}{\frac{\partial y_T}{\partial V_s} + \frac{\partial y_T}{\partial L_R} \left[\frac{-\frac{\partial x_B}{\partial V_s}}{\frac{\partial x_B}{\partial L_R}} \right]} \quad (20.34a)$$

$$\lambda_3 = \frac{\left. \frac{\Delta x_B}{\Delta L_R} \right|_{\text{constant } V_s}}{\left. \frac{\Delta x_B}{\Delta L_R} \right|_{\text{constant } y_T}} = \frac{\frac{\partial x_B}{\partial L_R}}{\frac{\partial x_B}{\partial L_R} + \frac{\partial x_B}{\partial V_s} \left[\frac{-\frac{\partial y_T}{\partial L_R}}{\frac{\partial y_T}{\partial V_s}} \right]} \quad (20.34b)$$

$$\lambda_4 = \frac{\left. \frac{\Delta x_B}{\Delta V_s} \right|_{\text{constant } L_R}}{\left. \frac{\Delta x_B}{\Delta V_s} \right|_{\text{constant } y_T}} = \frac{\frac{\partial x_B}{\partial V_s}}{\frac{\partial x_B}{\partial V_s} + \frac{\partial x_B}{\partial L_R} \left[\begin{array}{c} -\frac{\partial y_T}{\partial V_s} \\ \frac{\partial y_T}{\partial L_R} \end{array} \right]} \quad (20.34c)$$

Further, the theory says that:

$$\lambda_2 = 1 - \lambda_1$$

$$\lambda_3 = 1 - \lambda_1$$

$$\lambda_4 = \lambda_1$$

We can see that each column and each row must add up to 1. If $\lambda_1 \cong \lambda_2 \cong \lambda_3 \cong \lambda_4$, interaction is severe.

Relative-Gain Matrix for Binary Column—II

This scheme is the one that Shinskey⁵ terms “material-balance” control:

1. Condensate receiver level controlled by reflux.
2. Top composition controlled by distillate.
3. Bottom composition controlled by boilup (usually steam flow).
4. Base level controlled by bottom product withdrawal.

The two starting equations are:

$$\Delta y_T = \frac{\partial y_T}{\partial D} \Delta D + \frac{\partial y_T}{\partial V_s} \Delta V_s \quad (20.35)$$

and

$$\Delta x_B = \frac{\partial x_B}{\partial D} \Delta D + \frac{\partial x_B}{\partial V_s} \Delta V_s \quad (20.36)$$

so:

$$\Lambda = \begin{bmatrix} \lambda_1 & \lambda_2 \\ \lambda_3 & \lambda_4 \end{bmatrix} \quad (20.37)$$

Here:

$$\lambda_1 = \frac{\left. \frac{\Delta y_T}{\Delta D} \right|_{\text{constant } V_s}}{\left. \frac{\Delta y_T}{\Delta D} \right|_{\text{constant } x_B}} \quad (20.38)$$

By a procedure similar to that of "Binary Column—I" we find:

$$\lambda_1 = \frac{\frac{\partial y_T}{\partial D}}{\frac{\partial y_T}{\partial D} + \frac{\partial y_T}{\partial V_s} \left[\frac{-\frac{\partial x_B}{\partial D}}{\frac{\partial x_B}{\partial V_s}} \right]} \quad (20.39)$$

Next:

$$\lambda_2 = \frac{\left. \frac{\Delta y_T}{\Delta V_s} \right|_{\text{constant } D}}{\left. \frac{\Delta y_T}{\Delta V_s} \right|_{\text{constant } x_B}} = \frac{\frac{\partial y_T}{\partial V_s}}{\frac{\partial y_T}{\partial V_s} + \frac{\partial y_T}{\partial D} \left[\frac{-\frac{\partial x_B}{\partial V_s}}{\frac{\partial x_B}{\partial D}} \right]} \quad (20.39a)$$

$$\lambda_3 = \frac{\left. \frac{\Delta x_B}{\Delta D} \right|_{\text{constant } V_s}}{\left. \frac{\Delta x_B}{\Delta D} \right|_{\text{constant } y_T}} = \frac{\frac{\partial x_B}{\partial D}}{\frac{\partial x_B}{\partial D} + \frac{\partial x_B}{\partial V_s} \left[\frac{-\frac{\partial y_T}{\partial D}}{\frac{\partial y_T}{\partial V_s}} \right]} \quad (20.39b)$$

$$\lambda_4 = \frac{\left. \frac{\Delta x_B}{\Delta V_s} \right|_{\text{constant } D}}{\left. \frac{\Delta x_B}{\Delta V_s} \right|_{\text{constant } y_T}} = \frac{\frac{\partial x_B}{\partial V_s}}{\frac{\partial x_B}{\partial V_s} + \frac{\partial x_B}{\partial D} \left[\frac{-\frac{\partial y_T}{\partial V_s}}{\frac{\partial y_T}{\partial D}} \right]} \quad (20.39c)$$

Again:

$$\lambda_2 = 1 - \lambda_1$$

$$\lambda_3 = 1 - \lambda_1$$

$$\lambda_4 = \lambda_1$$

Relative-Gain Matrix for Binary Column—III

The last scheme we consider here has the following features:

1. Condensate receiver level controlled by distillate flow.
2. Top composition controlled by manipulation of L_R/D .

3. Bottom composition controlled by boilup (usually steam flow).
4. Column-base level controlled by bottom-product withdrawal.

First let:

$$\frac{L_R}{D} = R \quad (20.40)$$

Then:

$$\Delta y_T = \frac{\partial y_T}{\partial R} \Delta R + \frac{\partial y_T}{\partial V_s} \Delta V_s \quad (20.41)$$

$$\Delta x_B = \frac{\partial x_B}{\partial R} \Delta R + \frac{\partial x_B}{\partial V_s} \Delta V_s \quad (20.42)$$

whence, once more:

$$\Lambda = \begin{bmatrix} \lambda_1 & \lambda_2 \\ \lambda_3 & \lambda_4 \end{bmatrix} \quad (20.43)$$

Here:

$$\lambda_1 = \frac{\left. \frac{\Delta y_T}{\Delta R} \right|_{\text{constant } V_s}}{\left. \frac{\Delta y_T}{\Delta R} \right|_{\text{constant } x_B}} \quad (20.44)$$

By a procedure similar to that used in the two previous sections, we find:

$$\lambda_1 = \frac{\frac{\partial y_T}{\partial R}}{\frac{\partial y_T}{\partial R} + \frac{\partial y_T}{\partial V_s} \left[\frac{-\frac{\partial x_B}{\partial R}}{\frac{\partial x_B}{\partial V_s}} \right]} \quad (20.45)$$

Next:

$$\lambda_2 = \frac{\left. \frac{\Delta y_T}{\Delta V_s} \right|_{\text{constant } R}}{\left. \frac{\Delta y_T}{\Delta V_s} \right|_{\text{constant } x_B}} = \frac{\frac{\partial y_T}{\partial V_s}}{\frac{\partial y_T}{\partial V_s} + \frac{\partial y_T}{\partial R} \left[\frac{-\frac{\partial x_B}{\partial V_s}}{\frac{\partial x_B}{\partial R}} \right]} \quad (20.45a)$$

$$\lambda_3 = \frac{\left. \frac{\Delta x_B}{\Delta R} \right|_{\text{constant } V_s}}{\left. \frac{\Delta x_B}{\Delta R} \right|_{\text{constant } y_T}} = \frac{\frac{\partial x_B}{\partial R}}{\frac{\partial x_B}{\partial R} + \frac{\partial x_B}{\partial V_s} \left[\frac{-\frac{\partial y_T}{\partial R}}{\frac{\partial y_T}{\partial V_s}} \right]} \quad (20.45b)$$

$$\lambda_4 = \frac{\left. \frac{\Delta x_B}{\Delta R_s} \right|_{\text{constant } R}}{\left. \frac{\Delta x_B}{\Delta V_s} \right|_{\text{constant } y_T}} = \frac{\frac{\partial x_B}{\partial V_s}}{\frac{\partial x_B}{\partial V_s} + \frac{\partial x_B}{\partial R} \left[\frac{-\frac{\partial y_T}{\partial V_s}}{\frac{\partial y_T}{\partial R}} \right]} \quad (20.45c)$$

Again:

$$\lambda_4 = \lambda_1$$

$$\lambda_2 = \lambda_3 = 1 - \lambda_1$$

Example. As an example let us consider a binary system with low boiler *A* and high boiler *B*. The design specification for y_T is 0.985710 mFA and for x_B is 0.000165 mFA. An HP-41C program was used to calculate curves of L_R/D versus y_T (Figure 20.8), L_R/D versus x_B (Figure 20.9), V versus y_T (Figure 20.10), and V versus x_B (Figure 20.11). At the specified operating conditions, slopes were drawn on each of these curves to obtain the four gains. Originally these slopes were drawn by eye, but this led to serious errors in calculating λ 's. The final results, determined by digital differentiation, are as follows:

$$\frac{\partial y_T}{\partial R} = 0.215 \frac{\text{mFA}}{\left(\frac{\text{m/hr}}{\text{m/hr}} \right)}$$

$$\frac{\partial x_B}{\partial R} = 1.47 \frac{\text{mFA}}{\left(\frac{\text{m/hr}}{\text{m/hr}} \right)}$$

$$\frac{\partial y_T}{\partial V_s} = -0.0000422 \frac{\text{mFA}}{\text{m/hr}}$$

$$\frac{\partial x_B}{\partial V_s} = -0.000483 \frac{\text{mFA}}{\text{m/hr}}$$

From equation (20.45) we find:

$$\lambda_1 = 2.47$$

Hence:

$$\lambda_4 = 2.47$$

and

$$\lambda_2 = -1.47 = \lambda_3$$

so:

$$\Lambda = \begin{bmatrix} 2.47 & -1.47 \\ -1.47 & 2.47 \end{bmatrix}$$

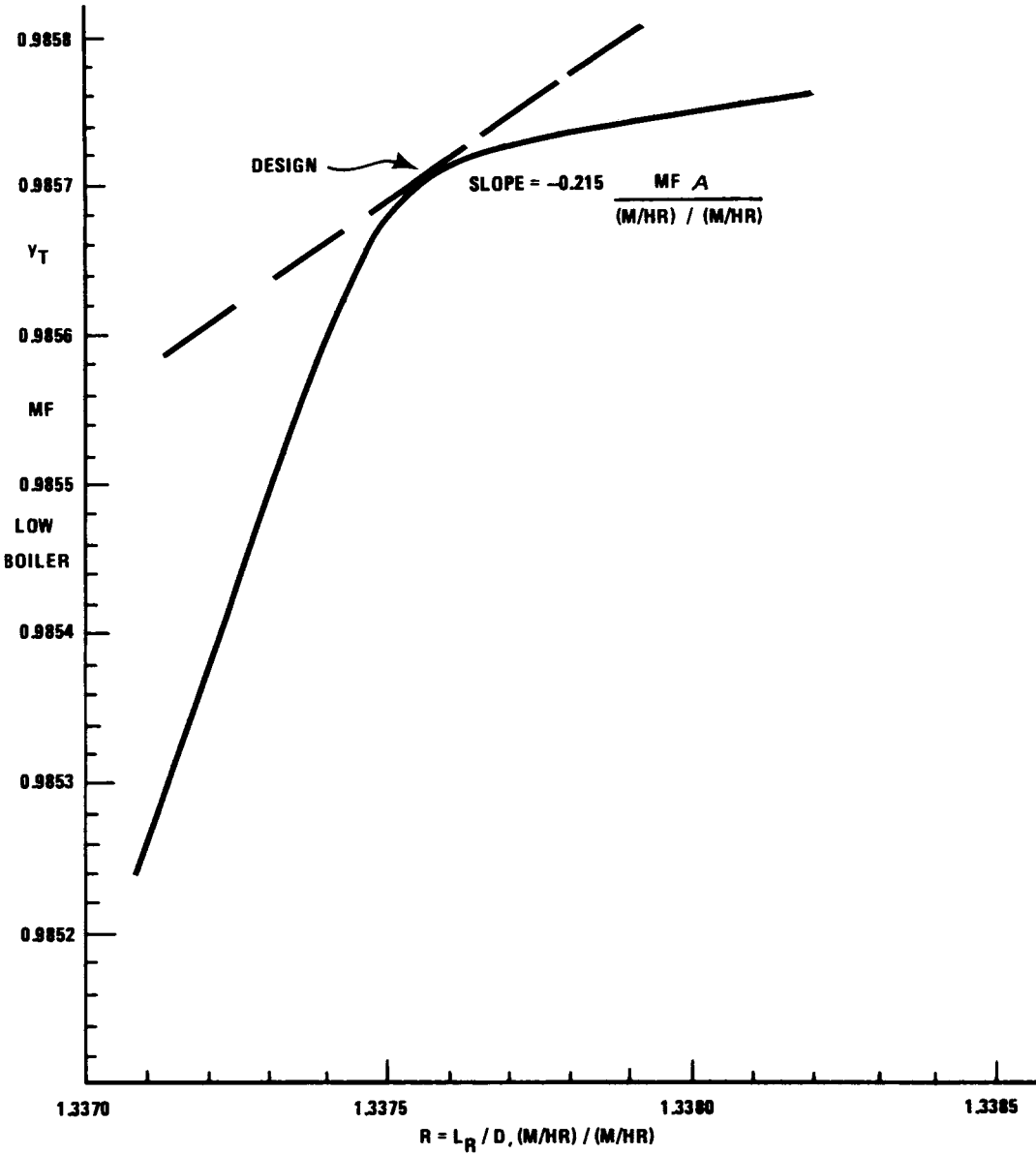


FIGURE 20.8
 y_T vs. R

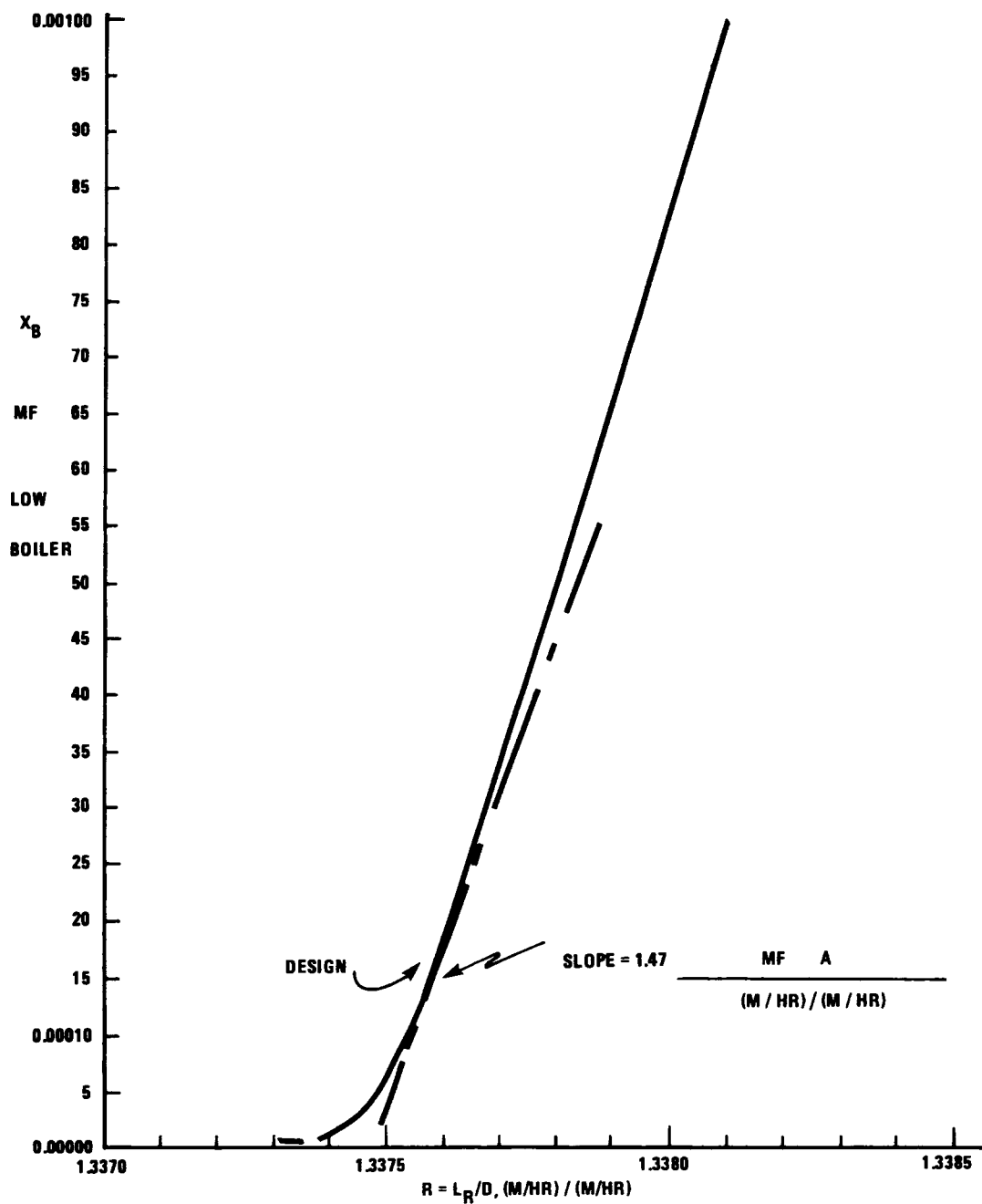


FIGURE 20.9
 X_B vs. R

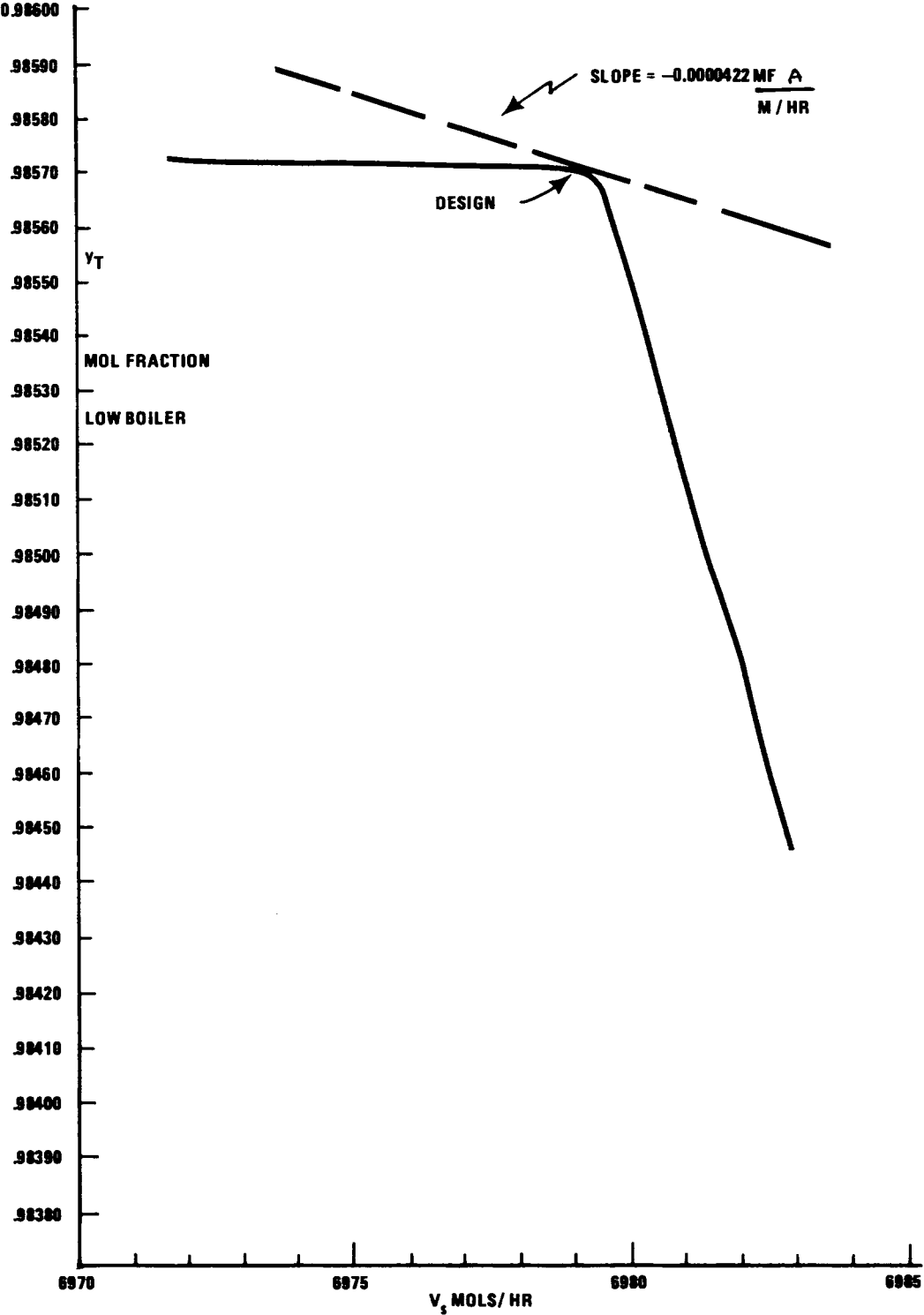


FIGURE 20.10
 y_T vs. V_s

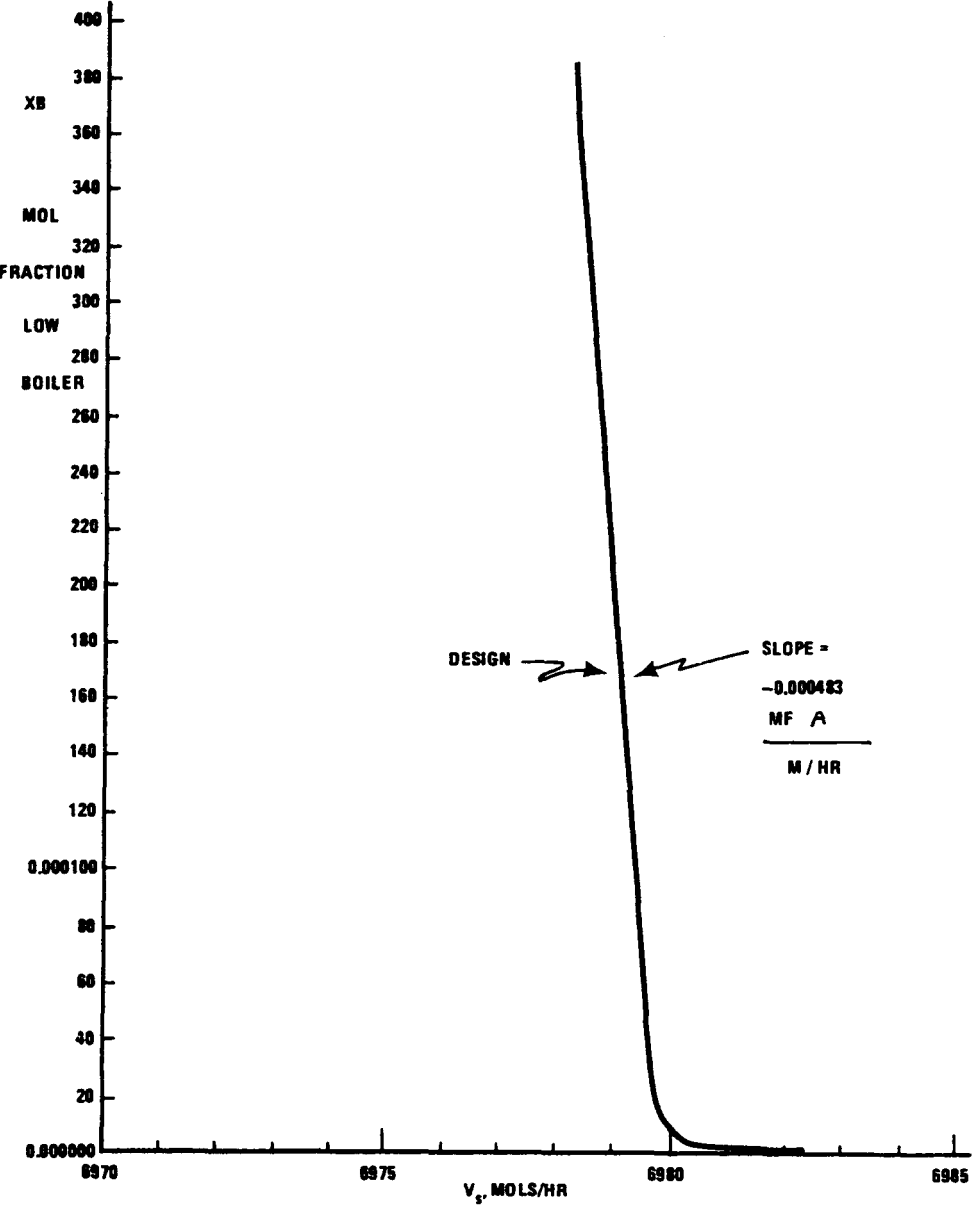


FIGURE 20.11
 X_B vs. V_s

Shinskey⁵ has derived simplified equations for column gains based on what he terms a "separation factor," S . For this system his equation for λ_1 is:

$$\lambda_1 = \left[\frac{1}{1 + \frac{(y_T - z_F) x_B (1 - x_B)}{(z_F - x_B) y_T (1 - y_T)}} \right] \times \left[1 + \frac{y_T - x_B}{y_T (1 - y_T) \ln S} \right] \quad (20.46)$$

where

$$S = \frac{y_T (1 - x_B)}{x_B (1 - y_T)} \quad (20.47)$$

For the illustrative problem, $z_F = 0.486933$ mfA, so solution of equations (20.46) and (20.47) gives $\lambda_1 = 6.33$. Then:

$$\Lambda = \begin{array}{cc} R & V \\ \left[\begin{array}{cc} 6.33 & -5.33 \\ -5.33 & 6.33 \end{array} \right] \end{array}$$

These results are clearly quite different from those obtained by the more rigorous method. This may be partly due to the high terminal purities. In making comparisons of the two methods with other columns, we found that for terminal purities of 95 mol percent or less, the two methods gave comparable results. The range of validity of Shinskey's separation factor unfortunately has never been published. In view of the ease of calculating steady-state column gains rigorously with programmable calculators or small computers, we recommend the latter approach.

20.6 COMPOSITION MEASUREMENT LOCATION

In the distillation literature, considerable attention has been devoted to selecting measurement locations for composition control. One might think offhand that to control terminal composition, one should measure compositions at the column top and bottom. This turns out indeed to be the case. For overhead composition control, column-located measurements should be as close to the top tray as possible, but not on the top tray where there is usually poor mixing. For some columns overhead vapor or condensate composition measurements may be feasible. For column bottom composition control, the measurement should be located in the liquid line to the reboiler or in the vapor space below the bottom tray. Measurements on any interior tray of the column should be avoided if at all possible for two reasons: (1) the relationship between composition on any interior tray and composition at the top or bottom of the column is ambiguous, and (2) it has been shown that the closer the measurement to the point of manipulation, the better is the control.

The last statement warrants some emphasis in view of the various schemes found in the literature for selecting a "control" tray. Two common ones are (1) that tray where the temperature change per tray is a maximum, and (2) that tray where the change in temperature per change in overhead (or bottoms) composition is a maximum.¹⁶ None of these schemes has any validity for feedback control in binary systems unless it is physically impossible to make a measurement at the ends of the column, and this is seldom the case with today's measurement technology. Uitti¹⁶ has shown that for an ideal binary system, the control of composition at a point above the feed tray and below the top tray leads to an odd result: for an increase in the feed concentration of the more volatile component, the concentration of the same component in the overhead decreases. For the same system, Uitti showed that by maintaining a constant reflux-to-feed ratio, he got fairly good control of overhead composition for feed composition changes. This was true even though he provided no feedforward compensation for these changes. Furthermore, an increase in feed concentration of the more volatile component led to an increase in overhead concentration of the same component. More recently Wood¹⁷ has shown that with the aid of a digital computer one may quickly calculate for a given control tray the steady-state variations in terminal composition for a specified change in feed composition.

Since holding interior tray temperatures constant does not hold terminal compositions constant in the face of feed-composition changes, it has occurred to some authors that one should vary interior tray temperatures as feed composition changes. Luyben,¹⁸ for example, has suggested a feedforward scheme for changing temperature controller set points.

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21

Sampled-Data Control of Distillation Columns

21.1 INTRODUCTION

Sampled-data control defines control action that is executed at discrete time intervals rather than continuously. An example is a composition control loop in which a gas chromatograph analyzer provides the composition measurement at 10-minute intervals. The controller output changes each time it receives a measurement, and then holds that value until a new measurement is received. Digital computers are ideally suited to this type of control because they operate at discrete intervals. Microprocessor-based control systems are also discrete devices but, because they are designed to replace analog equipment, their sampling time is less than 1 second, making them almost continuous devices.

Besides being used in gas chromatographic loops, sampled-data control is also useful in loops containing significant dead time, and where process gains are small and time constants are very large. Tray temperature control is one example where these guidelines apply. For large columns time constants in the range of one-half to one hour are not unusual. Employing continuous control is not necessary because the composition or temperature responds slowly to upsets.

Most sampled-data control systems employ discrete versions of PI and PID control algorithms although computers are certainly not limited to only these types. Special-purpose algorithms can be constructed in the software to deal with the multivariable, nonlinear nature of a distillation column. Adaptive control, for example, updates the parameters in control algorithms and sampling rates to compensate for nonlinearities in the process.¹ Optimal control is a

technique in which the parameters in the control algorithm are determined so that the process operates to minimize or maximize some index of performance.²⁻⁶ Another kind of optimal control, called “minimal prototype” control, selects the parameters to cause the process to line out smoothly in the minimum time following a load disturbance or a change in set point.⁷⁻¹⁰

Much still needs to be done to bring some of the new, modern control theories to the place where they are convenient to implement. Control engineers meanwhile have devised simpler methods to handle nonlinearities and the multivariable nature of a distillation column. Interaction compensation is a case in point. A change in one manipulative variable, for example, not only affects the variable it is controlling, but also may affect controlled variables in other loops. Therefore, interaction compensation attempts to eliminate the effect one manipulative variable has on other controlled variables. The majority of articles written about interaction compensation have been for continuous control (Chapter 20), while Fitzpatrick¹¹ and Shunta¹² designed sampled-data interaction compensators. Override control allows a valve to be manipulated by several different controllers, one at a time, according to a built-in logic that selects the controller to meet the most urgent need of the process (Chapter 9). Feedforward control is yet another type of multivariable control. A measured load disturbance is used to adjust the manipulative variable before feedback control occurs, to cancel the effect of the disturbance on the controlled variable (Chapter 20). Some simple adaptive control schemes have also been used. Adaptive sampling rates and gain tuning are typical examples.^{13,14} This chapter discusses the application of some of these concepts to sampled-data control of a distillation column.

21.2 CONTROL ALGORITHMS

Control algorithms have been the subject of numerous papers over the years and are still evolving today. The types of algorithms people have proposed range from simple proportional and discretized analog types (PI and PID) to highly complex optimal control algorithms. It is worthwhile here to review some of the more common types, their forms, and how their parameters are selected. More thorough discussions on this can be found in several texts on digital control.^{15,16}

Analog Types

The most common algorithms in use today are discretized PI and PID algorithms in which the continuous functions of integration and differentiation are approximated by numerical methods. A PID controller that is approximated by simple rectangular integration is:

$$M_n = K_c \left[e_n + \frac{T}{\tau_R} \sum_{k=1}^n e_k + \frac{\tau_D}{T} (e_n - e_{n-1}) \right] + B \quad (21.1)$$

This is the “position” form because the output M_n is the absolute value of the valve signal or the set point in supervisory (cascade) control. The bias term B is to initialize the output when the algorithm is put in service. An alternate approximation is called the “velocity” form whose output is the change in valve position or set point.

$$\Delta M_n = K_c \left[(e_n - e_{n-1}) + \frac{T}{\tau_R} e_n + \frac{\tau_D}{T} (e_n - 2e_{n-1} + e_{n-2}) \right] \quad (21.2)$$

The velocity form does not have to be initialized. Its output changes the position of a stepping motor or other integrating device that, in turn, positions the valve or set point. Because these algorithms approximate the continuous ones, their performance deteriorates as the sampling period T increases.¹⁷

The crux in implementing these algorithms is to select the parameters K_c , τ_R , τ_D , and T to achieve satisfactory control. This, of course, requires knowing what the actual process dynamics are. Controller tuning on line determines the dynamics directly by manually changing the set point and observing the process transient response. Controller tuning off line needs a dynamic model of the process. From the dynamics, the controller parameters are selected to meet certain performance criteria. Some criteria are similar to those for continuous systems, for example, damping ratio, closed-loop M_p , and so forth. Others select the parameters to minimize an index of performance such as, for example, integral-error-squared.^{18–20}

Sampled-Data Algorithms

An alternative to discrete PI or PID algorithms is one that is determined by sampled-data techniques using the z -transformation. This algorithm does not have parameters K_c , τ_R , and τ_D , but is expressed as a ratio of polynomials in powers of Z whose coefficients are specified to achieve a certain response. Some criteria for selecting coefficients are like the methods described in the previous section while others select the coefficients to obtain a specified type of closed-loop response.^{21–28} For example, Dahlin’s method²⁸ specifies the response to a step change in set point to be a first-order lag with dead time. The “minimal prototype” sampled-data algorithm is a type of optimal control in which the output is specified to reach set point, that is, zero error, in the fastest time without exhibiting oscillations.

The methods discussed thus far for selecting parameters apply to single-loop control that is limited to a linear region, usually ± 10 – 15 percent about the initial steady state. Provision should be made to adapt the algorithm to process nonlinearities, that is, changes in process gains and time constants as the operating conditions change. Gallier¹³ discusses a typical method. In addition, if process interactions are significant, designing for single-loop control may be inadequate and some sort of interaction compensation will be required.

21.3 SERVO AND REGULATOR CONTROL

Having discussed some general points about sampled-data control techniques and algorithms, we now look specifically at how sampled-data control is applied to a distillation column. Fundamentally we are interested in achieving good control in the face of set-point changes and load disturbances—traditionally called servo and regulator control.

Let us consider a top- or bottom-product composition-control loop. A common approach for binary systems is to control the temperature on some tray in the rectifying or stripping section in the absence of an on-line product analyzer. A typical loop is shown in Figure 21.1 where steam flow is manipulated to control the temperature on a stripping tray. The temperature control loop is in the computer, and its output is the set point to the analog steam-flow controller. The performance requirements for this loop are that we get a smooth transition from one temperature to another when a set-point change is made, and that the temperature undergoes a tolerable deviation when a load disturbance

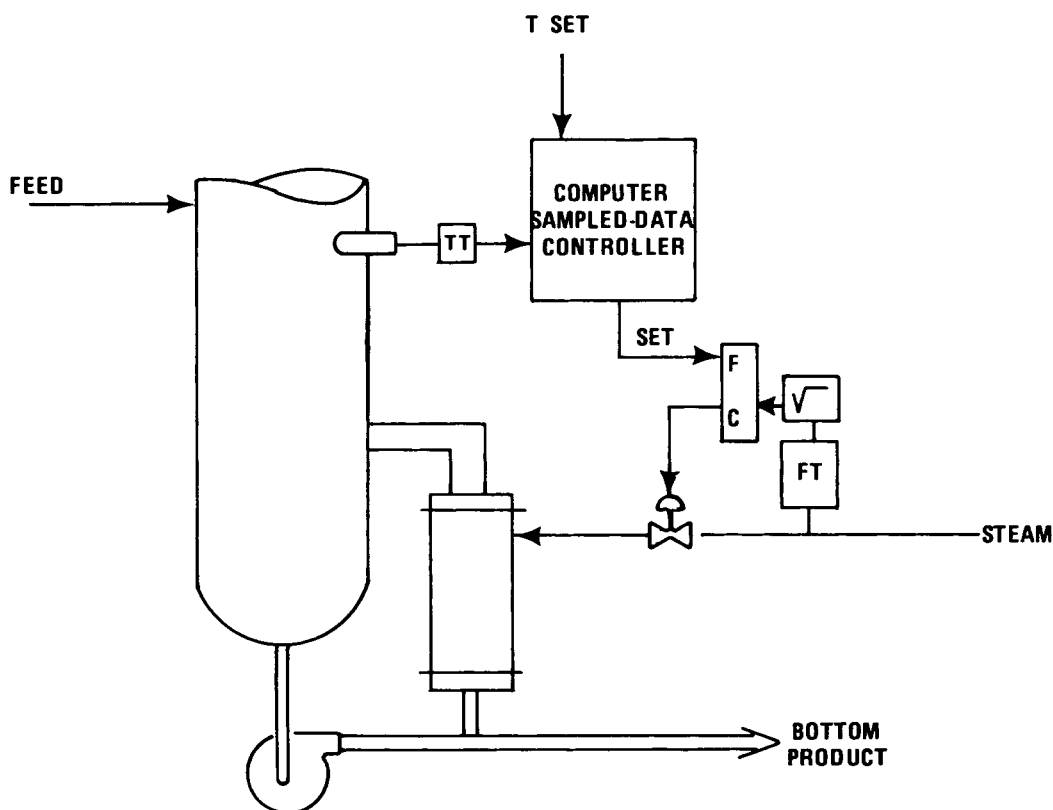


FIGURE 21.1
Sampled-data control

in feed rate or feed composition occurs. A set-point change requires a different response of the controlled variable than a load disturbance requires. Consequently a controller tuned for good response to a set-point change may be unsatisfactory for load disturbances. When a conventional discrete PI or PID algorithm is used as shown in Figure 21.2, the parameters can be specified for either good servo or regulator control, but not for both. Some have proposed that two separate sets of tuning parameters be stored in the computer. This approach requires additional computer logic and space.

Dual Algorithm

A better approach is to implement a “dual” sampled-data algorithm that is structured to handle both set-point and load disturbances simultaneously, as shown in Figure 21.3.²⁹ The $D_L(z)$ part of the algorithm is designed to achieve good regulator control and the $D_s(z)$ part to achieve good servo control. These algorithms are derived in the following manner.

The equation for the controlled variable in sampled notation is derived from Figure 21.3. We assume the dynamics of the continuous-flow controller are fast enough to be neglected.

$$C(z) = \frac{C^{\text{set}}(z)D_s(z)D_L(z)G_pH(z)}{1 + D_L(z)G_pH(z)} + \frac{G_L L(z)}{1 + D_L(z)G_pH(z)} \quad (21.3)$$

$G_pH(z)$ is the z -transform of the product of the process transfer function and zero-order hold. The output of the zero-order hold is the last value of the computer output, which is held constant until the next sample time. $G_L L(z)$ is the z -transform of the product of the load transfer function and the load variable. $D_L(z)$ is determined from equation (21.3) by setting $C^{\text{set}}(z)$ equal to zero.

$$D_L(z) = \frac{G_L L(z) - C(z)}{G_p H(z) C(z)} \quad (21.4)$$

$G_L L(z)$ and $C(z)$ have to be specified before $D_L(z)$ can be calculated. Therefore, some knowledge of the type of load is necessary—whether it is a step or ramp function. The time response of $C(z)$ to the load disturbance can be specified to meet any number of criteria as long as $D_L(z)$ is physically realizable. $C(z)$ is in the form of a series of negative powers of z . The power of z corresponds to the number of sample points following the disturbance. The coefficient of z is the deviation from the initial value of $C(z)$.

The set-point compensation part of the algorithm $D_s(z)$ is determined from equation (21.3) by setting $G_L L(z)$ equal to zero. This says that the load variable does not change.

$$D_s(z) = \frac{C(z)[1 + D_L(z)G_pH(z)]}{C^{\text{set}}(z)D_L(z)G_pH(z)} \quad (21.5)$$

The appropriate terms must be substituted into equation (21.5) to calculate $D_s(z)$. $D_L(z)$ has been calculated already from equation (21.4). $C^{\text{set}}(z)$ is the z -

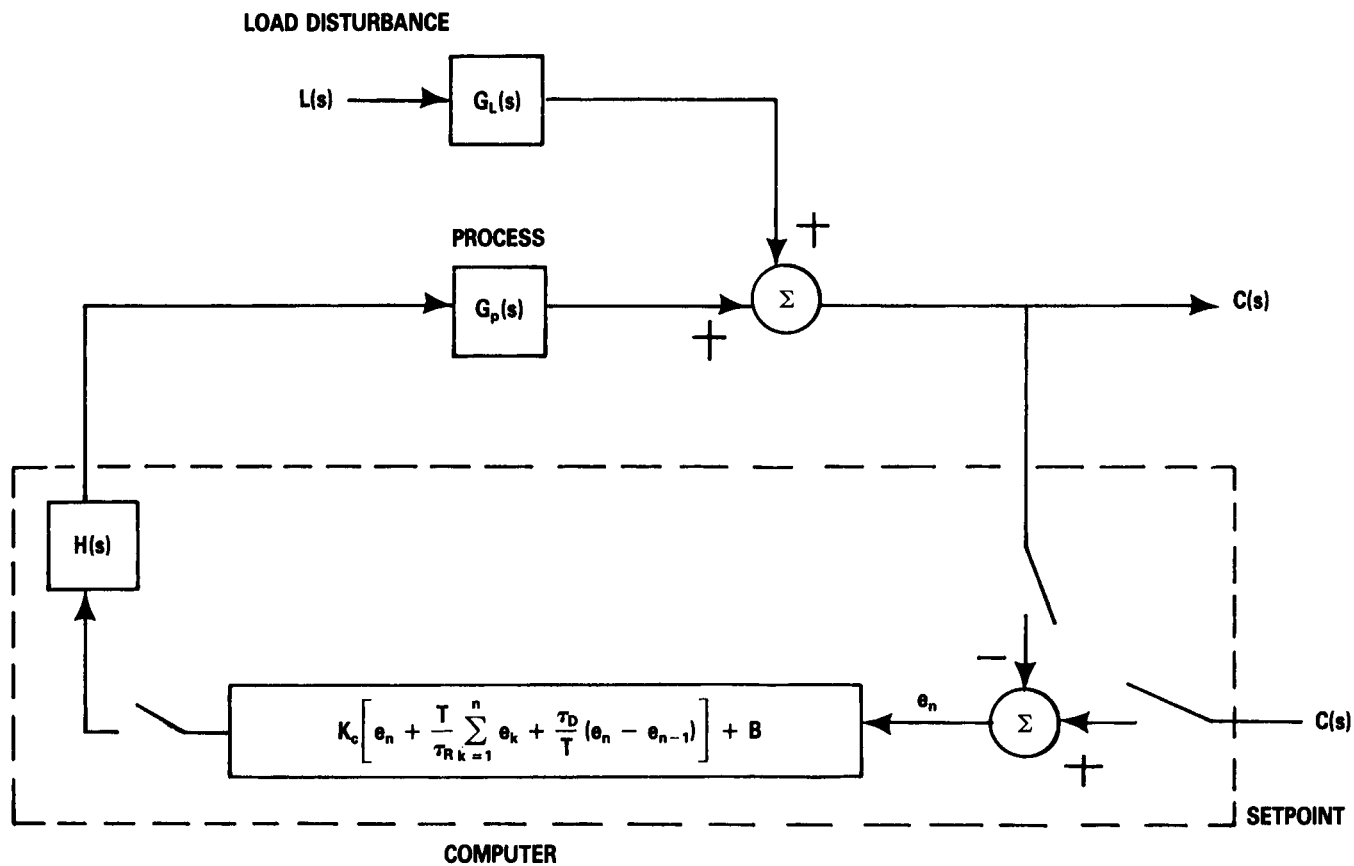


FIGURE 21.2
Discrete PID sampled-data control

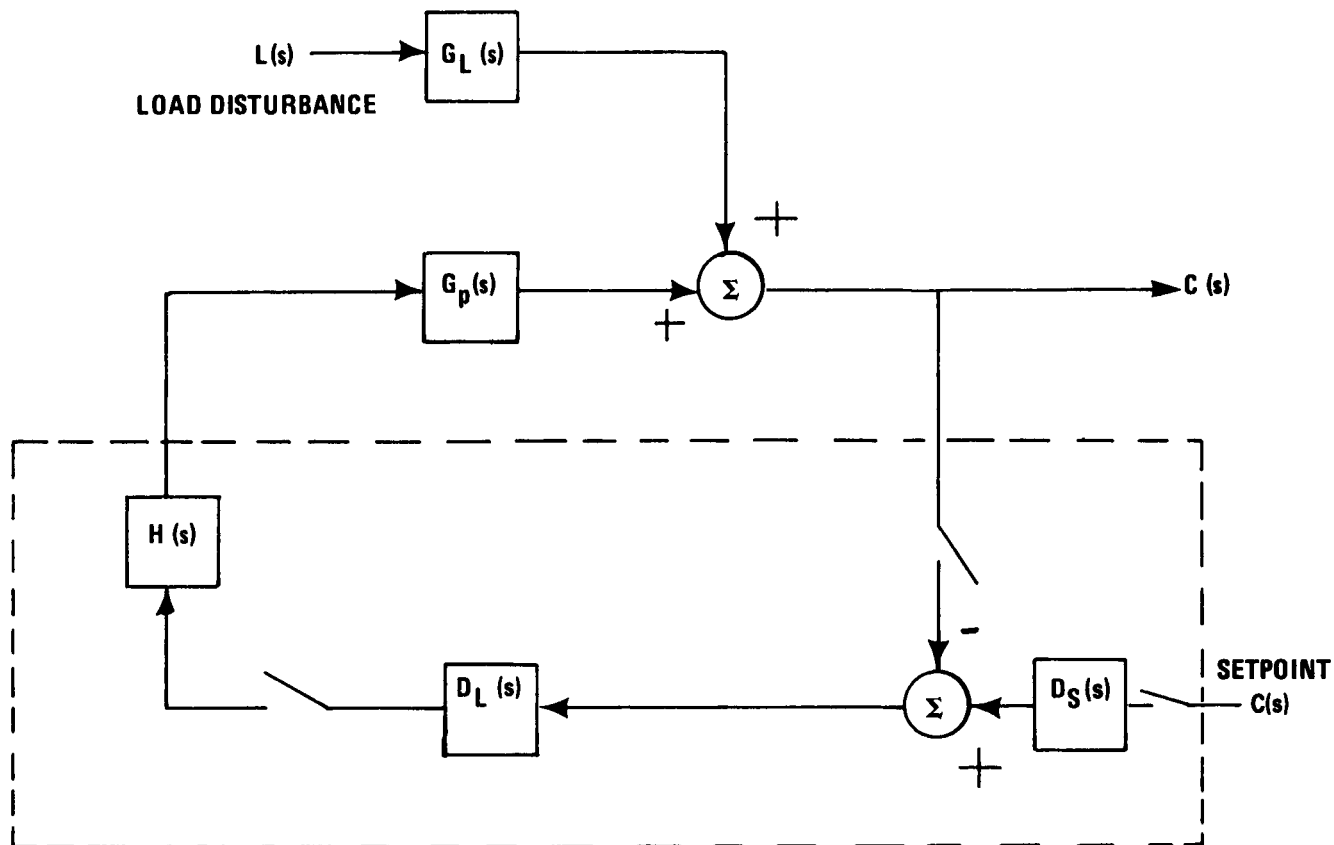


FIGURE 21.3
"Dual" sampled-data control

transform of the set-point change, typically a step function. $C(z)$ is the specification of the controlled variable response to the set-point change.

Numerical Example

Cox and Shunta³⁰ report the transfer functions for a simulated 20-tray nonlinear binary distillation column in which the second tray temperature is controlled by manipulating steam. Let us determine the dual algorithm for this information (Table 21.1) and compare its performance with that of a discretized PID controller. We use the minimal prototype criterion in selecting the closed-loop response. The minimal prototype criterion specifies that $C(z)$ must return to set point in the fastest time possible following a load disturbance without having oscillations. $C(z)$ therefore is specified according to equation (21.6) for a step change in the load variable, feed composition. We express $C(z)$ as the mole fraction of the more volatile component on tray 2 instead of temperature.

$$C(z) = 0.13 \Delta x_f z^{-1} (1 + 0.65 z^{-1}) \quad (21.6)$$

$C(z)$ is the numerator of $G_L L(z)$. Δx_f is the magnitude of the load change. Equation (21.6) says that $C(t)$ will have a deviation of $0.13 \Delta x_f$ from set point at the first sampling instant following the feed composition disturbance and a deviation of $0.0845 \Delta x_f$ two sampling instants past the disturbance. $C(t)$ will be at set point for the consequent sampling points. This assumes that the disturbance occurs immediately following a sampling instant. Therefore, $C(t)$ responds open loop for the sampling period. The appropriate terms from Table 21.1 and equation (21.6) can now be substituted into equation (21.4) to solve for $D_L(z)$.

$$D_L(z) = \frac{-730 (1 - 0.368 z^{-1})}{1 - z^{-1}} \quad (21.7)$$

Now let us solve for the servo part of the algorithm $D_s(z)$. The minimal prototype criterion says that $C(z)$ must reach the new set point as rapidly as possible without overshooting and having oscillations. $C(z)$ can be specified to reach set point in one sampling period for a first-order process, two sampling periods for a second-order process, and so on. This assumes that the process is linear and that we are not limited in the magnitude of the manipulative variable. $G_p(s)$ is almost first order, so specify $C(z)$ to reach the set point in one sampling period.

$$C(z) = \frac{\Delta C^{\text{set}} z^{-1}}{1 - z^{-1}} \quad (21.8)$$

ΔC^{set} is the magnitude of the set-point change. Substitute the appropriate terms in Table 21.1, and equations (21.7) and (21.8) into equation (21.5).

$$D_s(z) = \frac{0.5187}{1 - 0.48 z^{-1}} \quad (21.9)$$

TABLE 21.1
Transfer functions

LAPLACE TRANSFORM	Z-TRANSFORM
Process transfer function: $X_2(s) / V(s)$	
$G_p(s) = \frac{-0.281 (1.33s + 1)}{(12.5s + 1) (s + 1)}$	$G_p^H(z) = \frac{-0.0264z^{-1} (1 - 0.48z^{-1})}{(1 - 0.923z^{-1}) (1 - 0.368z^{-1})}$
Zero order hold:	
$H(s) = \frac{1 - e^{-Ts}}{s}$	
Load transfer function: $X_2(s) / X_F(s)$	
$G_L(s) = \frac{2.94e^{-0.4s}}{13.3s + 1}$	$G_L^L(z) = \frac{0.13 Lz^{-1} (1 + 0.65z^{-1})}{(1 - 0.928z^{-1}) (1 - z^{-1})}$
Load:	
$L(s) = \frac{\Delta L}{s}$	
Setpoint:	
$C(s) = \frac{\Delta C}{s}$	$C(z) = \frac{\Delta C}{1 - z^{-1}}$

The output of $D_i(z)$ in the time domain is a function of the present value of set point $C^{\text{set}}(t)$ and the last value of $D_i(t)$.

$$D_i(t) = 0.5187 C^{\text{set}}(t) + 0.48 D_i(t - T) \quad (21.10)$$

The output of $D_L(z)$ in the time domain is the required change in the manipulative variable.

$$D_L(t) = -730 [D_i(t) - C(t)] + 268 [D_i(t - T) - C(t - T)] + D_L(t - T) \quad (21.11)$$

Comparison of Dual and Discrete PID Algorithms

A nonlinear computer simulation³⁰ was used to compare the dual algorithm with a discretized PID algorithm for a set-point change and load disturbances in feed composition and feed rate. The PID settings were selected to meet a minimum integral of absolute error criterion for a set-point change as per Smith¹⁵ (page 176). The sampling period is 1 minute, approximately 10 percent of the process time constant. Figure 21.4a shows the response with the two controllers to a step change in set point from 0.057 to 0.01 mole fraction. Performance of the dual algorithm is slightly better than the PID algorithm.

Figure 21.4b shows the response to a step change in feed composition from 50 to 60 percent. The disturbance occurred immediately after sampling at $t = 0$. The response with the dual algorithm meets the specified response in equation (21.6) very well and is markedly better than with the PID controller.

A disadvantage of the minimal prototype criterion is that the algorithm is designed for one type of disturbance and may not be as good for others. This is illustrated for a disturbance in feed rate in Figure 21.4c. The feed rate is increased stepwise 30 percent and the algorithm derived for a feed-composition load is employed. The response with the dual algorithm exhibits some overshoot and oscillations, but is markedly better than the PID algorithm. In practice the load algorithm should be designed for the predominant disturbance.

It has thus been shown that the dual algorithm can handle both set-point and load disturbances satisfactorily without having to adjust the parameters. It is markedly better than a discrete PID algorithm tuned for set-point response.

21.4 FEEDFORWARD CONTROL

Feedforward control adjusts the manipulative variable as soon as a load disturbance is detected in an attempt to cancel its effect on the controlled variable. Sampled-data feedforward algorithms are designed readily from knowledge of the dynamics of the process.³¹ Feedforward control is best employed in conjunction with feedback control to correct for any offset due to inaccuracies in the process model. A typical feedforward/feedback sampled-data loop is

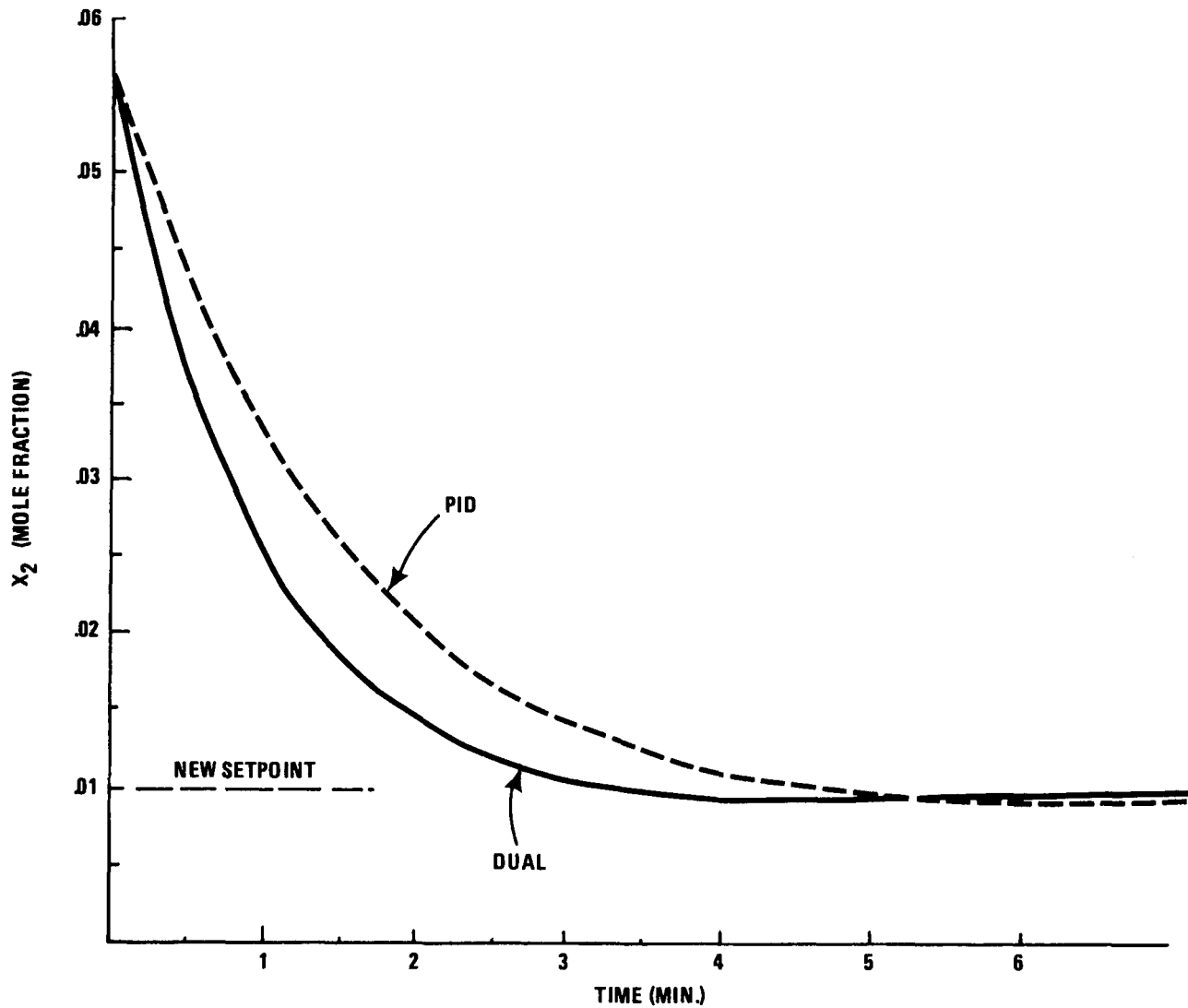


FIGURE 21.4a
"Dual" and PID set-point control

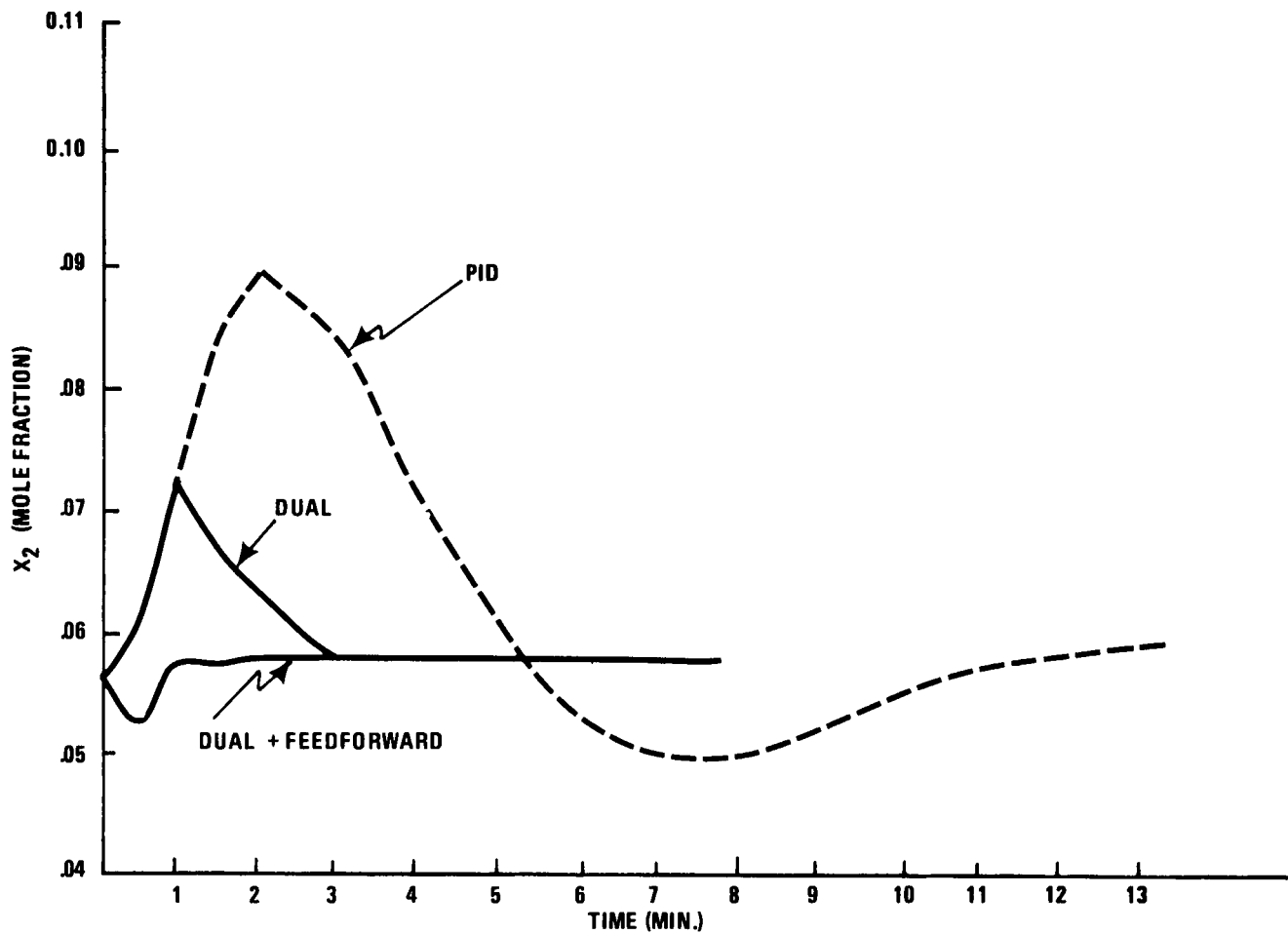


FIGURE 21.4b
"Dual" and PID control for feed composition disturbance

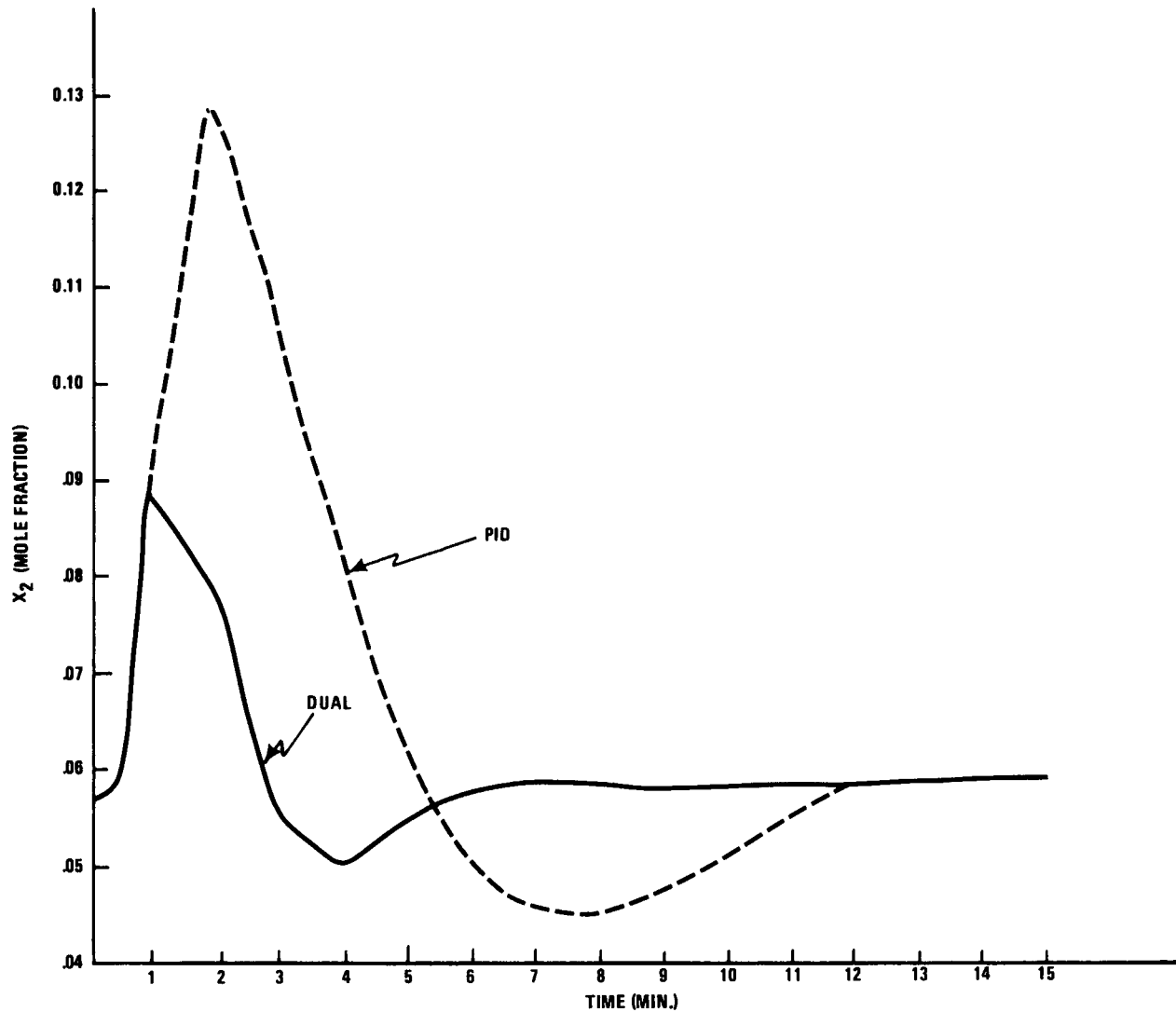


FIGURE 21.4c
"Dual" and PID control for feed rate disturbance

shown in Figure 21.5. Theoretically any number of feedforward algorithms $D_F(z)$ can be added. $D_F(z)$ is derived in the following manner.

A control-loop equation for $C(z)$ is written from Figure 21.5.

$$C(z) = \frac{L(z)D_F(z)G_pH(z) + G_L L(z)}{1 + D_L(z)G_pH(z)} \quad (21.12)$$

$C(z)$ is set equal to zero, which says that there will be zero error at each sampling point. $D_F(z)$ is solved from equation (21.12).

$$D_F(z) = \frac{-G_L L(z)}{L(z)G_pH(z)} \quad (21.13)$$

$D_F(z)$ can be determined from equation (21.13) once the nature of the load and the process model are known. Let us calculate the feedforward algorithm for a feed-composition disturbance from the information in Table 21.1 for a step-load change.

$$D_F(z) = \frac{49.3 (1 + 0.65 z^{-1})(1 - 0.368 z^{-1})}{(1 - 0.48 z^{-1})} \quad (21.14)$$

The time output $D_F(t)$ from equation (21.14) is the required feedforward change in the manipulative variable after a change in the feed composition is detected.

$$D_F(t) = 49.3 [\Delta x_f(t) + 0.28 \Delta x_f(t - T) - 0.24 \Delta x_f(t - 2T)] \\ + 0.48 D_F(t - T) \quad (21.15)$$

If the load form is not known exactly, $L(s)$ may be approximated by a staircase function $L^*(s)H(s)$.³¹ The asterisk denotes that $L(s)$ is a sampled variable. This simply says that the load is fictitiously sampled and the value is held constant for the sampling period. The numerator of equation (21.13) becomes:

$$G_L L(z) = L(z)G_L H(z) \quad (21.16)$$

Substitution of equation (21.16) into equation (21.13) gives a different form for $D_F(z)$.

$$D_F(z) = \frac{-G_L H(z)}{G_p H(z)} \quad (21.17)$$

Equation (21.17) can be solved for the feedforward algorithm without knowing the actual form of the load. If $L(s)$ is a step function, $D_F(z)$, from equation (21.17), will be equal to equation (21.13).

The nonlinear simulation was used to illustrate the closed-loop response of the controlled variable x_2 following a 30 percent increase in feed composition. The results are shown in Figure 21.4b with the feedback-only dual and PID algorithms. Control is immensely improved with the feedforward action. The slight deviation in x_2 with feedforward control is due to inaccuracies in the linear model and the long sampling time relative to the process dead time. The

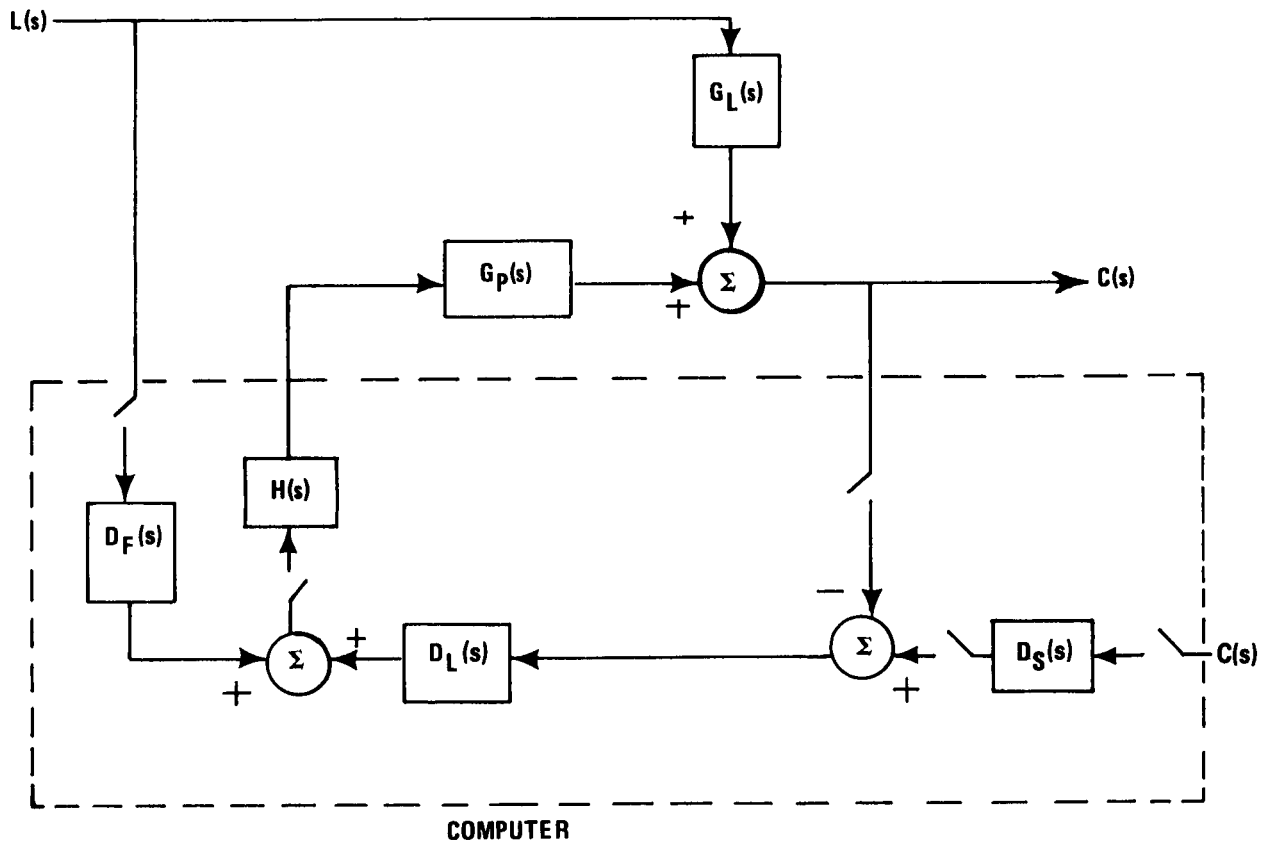


FIGURE 21.5
Sampled-data feed forward/feedback control loop

feedforward action initially disturbs the controlled variable in the direction opposite from the direction of the load disturbance so that the combined effects will cancel each other and x_2 will be at its set point at the following sampling points. The sampling time for this example should be equal to or less than the dead time in $G_L(s)$ so that the manipulation will occur when the effect of the load occurs.

21.5 INTERACTION COMPENSATION

Interaction occurs in multivariable systems when a change in a manipulative variable causes a deviation in more than one controlled variable.

A classic example is a distillation column in which a rectifying tray temperature is controlled by manipulating reflux, and stripping tray temperature by manipulating steam. A change in reflux causes a deviation in the stripping tray temperature. The steam flow is adjusted to compensate, and it causes a further deviation in the rectifying temperature. If the loops are tightly tuned, this interaction can result in unstable control. Chapter 20 discusses methods to eliminate interaction for analog controls. The purpose of this section is to illustrate how interaction compensators can be achieved by sampled-data algorithms for the control system discussed above. Basically the interaction compensators will function like feedforward algorithms with reflux and vapor boilup treated as load disturbances on the opposite controlled variables.¹² The compensators are used in conjunction with the dual feedback algorithms.

The control system is illustrated in Figure 21.6. The load and set-point algorithms D_{SS} , D_{LS} , D_{SR} , D_{LR} are designed according to Section 21.3. The interaction algorithms D_{IS} and D_{IR} are designed as follows. Consider first the stripping section control loop. The controlled variable X_s is given by equation (21.18):

$$X_s(z) = G_{IS}H(z)L_R(z) + G_{pS}H(z)V(z) \quad (21.18)$$

We desire zero change in $X_s(z)$ when a change is made in $L_R(z)$. Therefore, the necessary change to make in $V(z)$ is:

$$V(z) = - \frac{G_{IS}H(z)L_R(z)}{G_{pS}H(z)} \quad (21.19)$$

Equation (21.19) defines $D_{IS}(z)$:

$$D_{IS}(z) = \frac{V(z)}{L_R(z)} = - \frac{G_{IS}H(z)}{G_{pS}H(z)} \quad (21.20)$$

A similar calculation in the rectifying loop leads to $D_{IR}(z)$:

$$D_{IR}(z) = - \frac{G_{IR}H(z)}{G_{pR}H(z)} \quad (21.21)$$

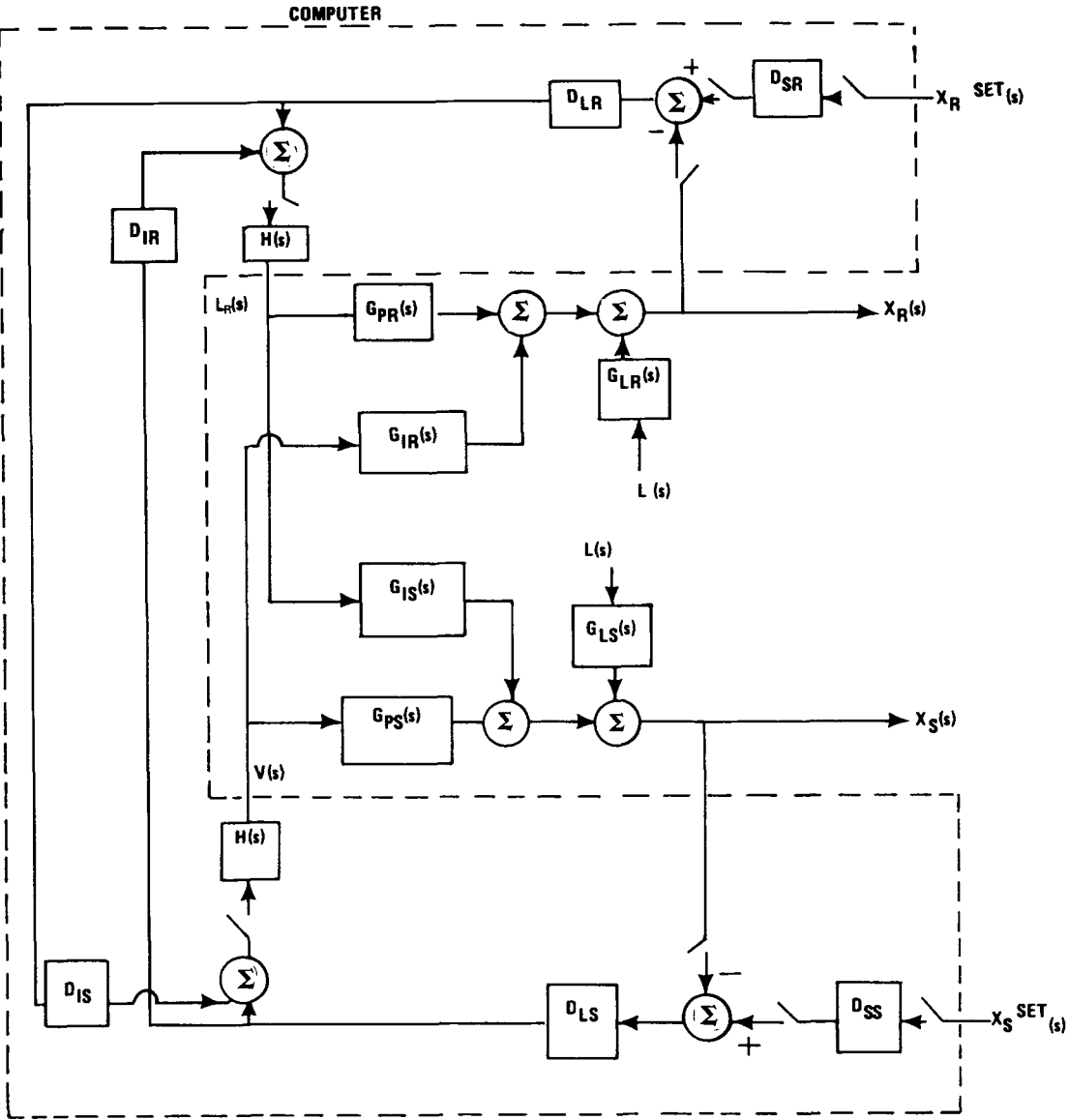


FIGURE 21.6
Interaction compensation

Thus, when a change in reflux or vapor boilup occurs, $D_{IS}(z)$ and $D_{IR}(z)$ make the appropriate change in the manipulative variable in the opposite loop to compensate for it.

Shunta¹² compares the results with and without sampled-data interaction compensators for controlling both ends of a 20-tray distillation column. Figures 21.7a and 21.7b show the closed-loop responses of composition on trays 2 and 18 for set-point changes and Figures 21.7c and 21.7d for a 10 percent change in feed composition. The improvement in control with interaction compensators is significant.

21.6 SAMPLED-DATA CONTROL FOR LOOPS WITH OVERRIDES

The use of overrides or protective controls as a way to deal with the multivariable nature of distillation-column control is discussed in Chapter 9 for analog controllers. The override control concept can be implemented readily in the computer as well. The advantage of computer implementation is the ease with which control schemes can be modified by simply reprogramming. Analog systems require changing hardware and rewiring. However, for safety considerations, hard-wired circuits are still preferred for critical overrides.

The principles of override control discussed in Chapter 9 suffice for sampled-data systems and will not be restated here. A tracking mechanism to prevent controller saturation or "windup" is needed for sampled-data systems as well as analog, but requires a different approach. This section, therefore, is devoted to the design of sampled-data algorithms that prevent windup.

Chapter 9 points out that an analog PI or PID controller saturates when an override takes over control of the valve unless the controller reset is made to track the valve signal in simple loops. The secondary variable measurement is the proper feedback for a primary controller in a cascade loop. Similarly, for sampled-data algorithms, the algorithm is structured to track an external feedback signal.³⁰

Tracking Dual Algorithm

We assume, for this discussion, that the overrides are implemented with analog hardware. This discussion, however, applies to a DDC loop as well. Figure 21.8 illustrates a conventional loop. We are again controlling the temperature in the stripping section of the distillation column by manipulating steam flow and using the dual algorithm. Overrides can enter the loop in the set-point path of the flow controller through the signal selector. The flow-control dynamics are very fast compared with the composition dynamics and, therefore, can be neglected for this analysis. With no means of tracking the output of the secondary variable (assumed equal to the output of the signal selector), the dual algorithm in Figure 21.8 saturates when an override occurs. Therefore, the algorithm must be restructured as in Figure 21.9. The secondary

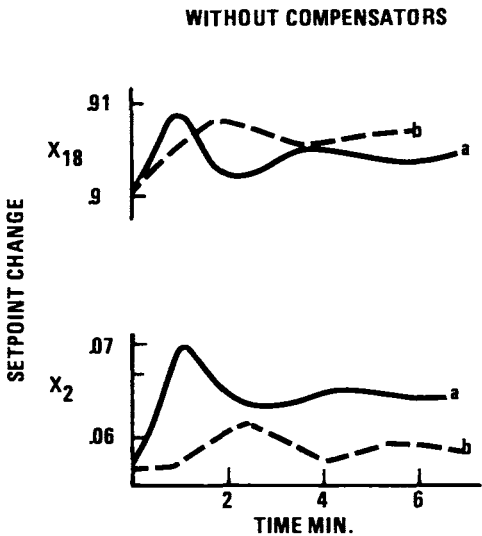


FIGURE 21.7a

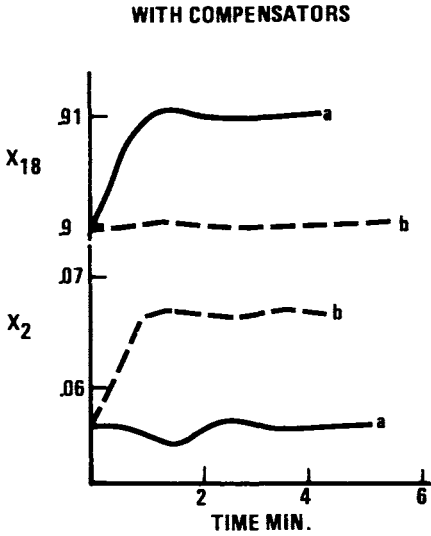


FIGURE 21.7b

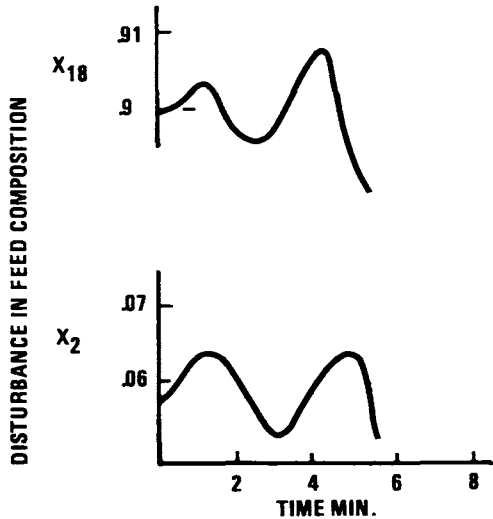


FIGURE 21.7c

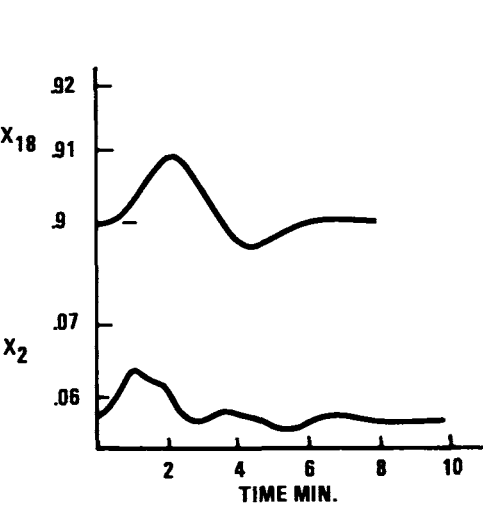


FIGURE 21.7d

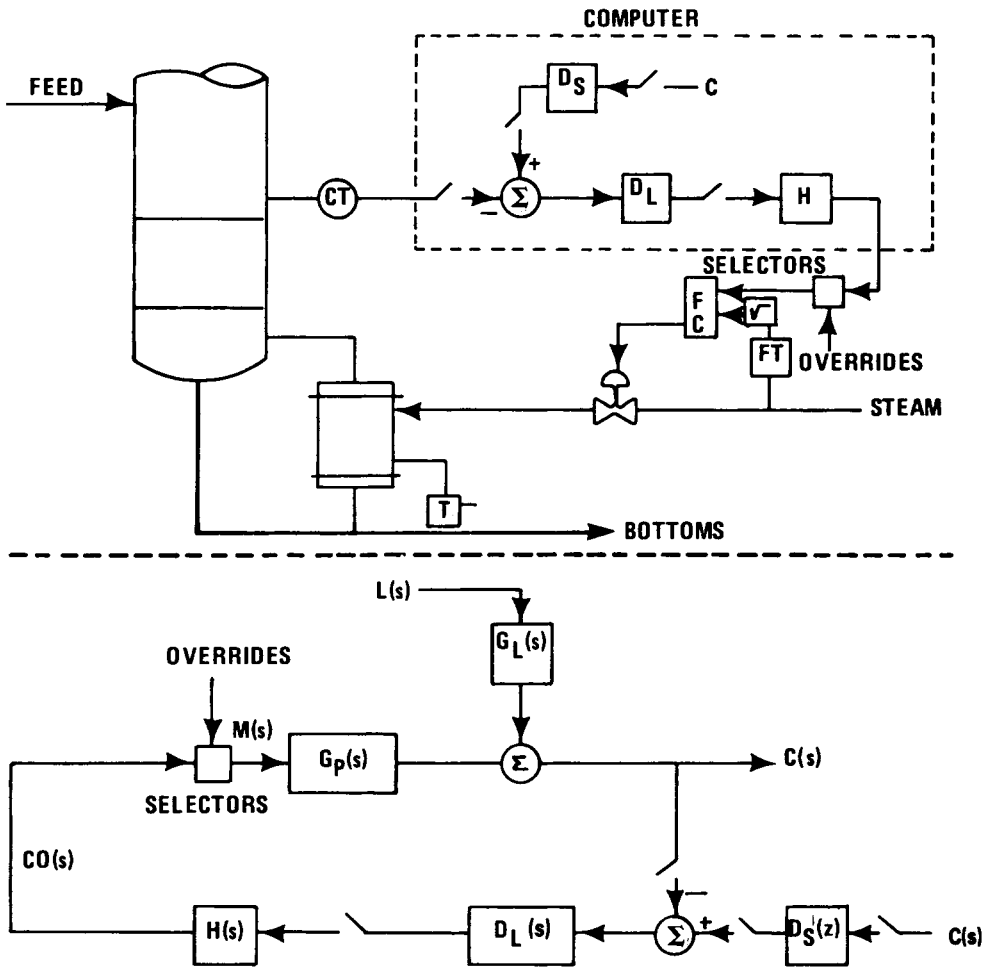


FIGURE 21.8
Conventional "dual" control in loop with overrides

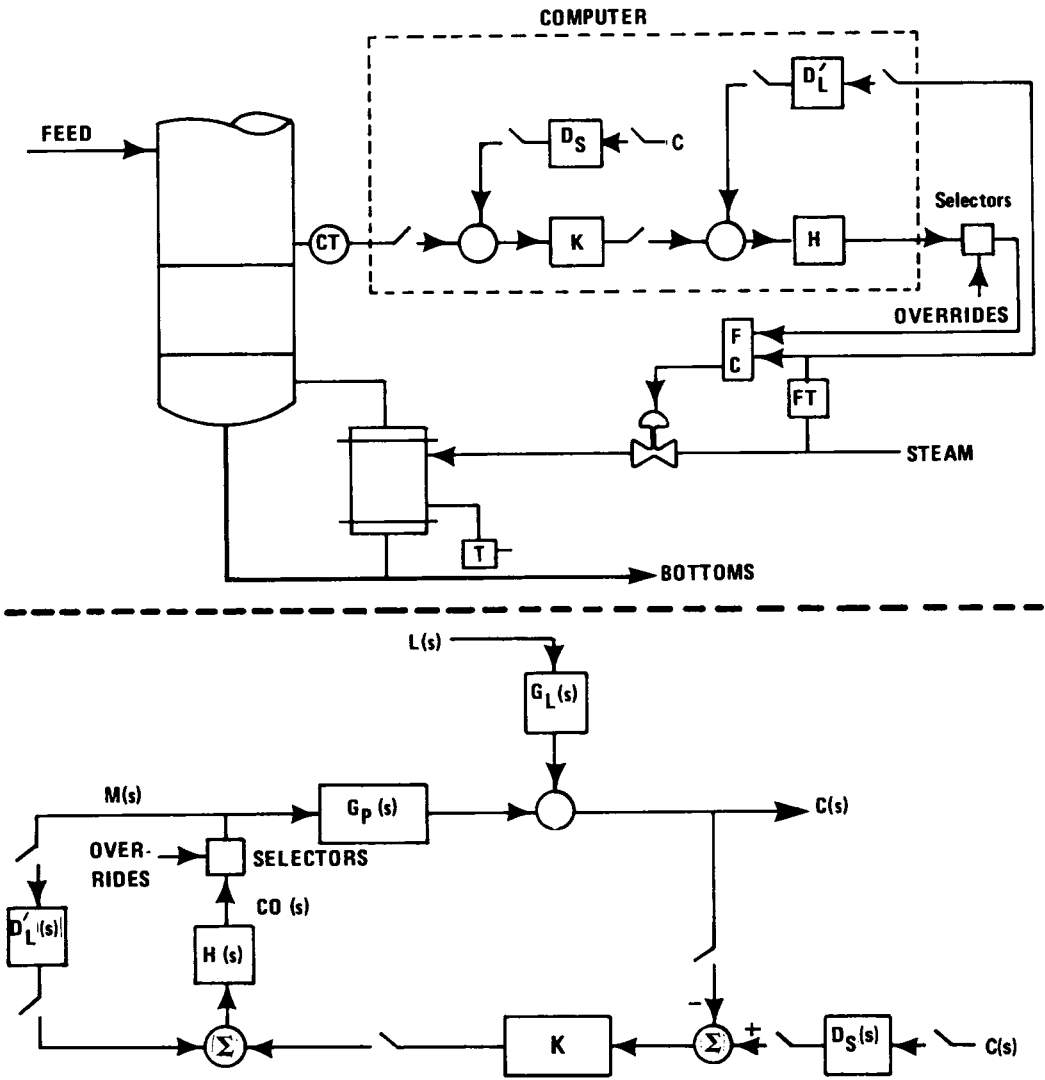


FIGURE 21.9
Tracking “dual” control in loop with overrides

variable is fed back to the computer at each sampling instant and is the input to the D'_L part of the algorithm. K is a static gain and its input is the difference between the set-point algorithm D_s and the controlled variable.

The algorithm D'_L is solved by first writing the loop equation. The overrides are assumed to be zero.

$$C(s) = [C^{\text{set}*}(s)D_s^*(s) - C^*(s)]KH(s)G_p(s) + M^*(s)D'_L(s)H(s)G_p(s) + L(s)G_L(s) \quad (21.22)$$

The asterisk (*) denotes a variable that has been sampled. The manipulative variable $M^*(s)$ is solved and substituted into equation (21.22).

$$\begin{aligned} M^*(s) &= M^*(s)D'_L(s) + [C^{\text{set}*}(s)D_s^*(s) - C^*(s)]K \\ &= \frac{[C^{\text{set}*}(s)D_s^{\text{set}*}(s) - C^*(s)]K}{1 - D'_L(s)} \end{aligned} \quad (21.23)$$

Therefore:

$$\begin{aligned} C(s) &= [C^{\text{set}*}(s)D_s^*(s) - C^*(s)]KH(s)G_p(s) \\ &+ \frac{[C^{\text{set}*}(s)D_s^*(s) - C^*(s)]KD'_L(s)H(s)G_p(s)}{1 - D'_L(s)} + L(s)G_L(s) \end{aligned} \quad (21.24)$$

Equation (21.24) is z -transformed and rearranged to solve for the controlled variable $C(z)$.

$$C(z) = \frac{C^{\text{set}}(z)D_s(z)G_pH(z)\left(\frac{K}{1 - D'_L(z)}\right) + G_L L(z)}{1 + G_pH(z)\left(\frac{K}{1 - D'_L(z)}\right)} \quad (21.25)$$

$D'_L(z)$ is solved by setting $C^{\text{set}}(z) = 0$.

$$D'_L(z) = \frac{G_L L(z) - G_pH(z)KC(z) - C(z)}{G_L L(z) - C(z)} \quad (21.26)$$

K is specified to make $D'_L(z)$ physically realizable, that is, so that z^0 terms in the numerator of $D'_L(z)$ are zero. $D'_L(z)$ is related to $D_L(z)$ by equation (21.27).

$$D'_L(z) = 1 - \frac{K}{D_L(z)} \quad (21.27)$$

This relationship holds true for any form of $D_L(z)$.

$D_s(z)$ is solved from equation (21.22) by setting $L(s) = 0$.

$$D_s(z) = \frac{C(z)}{C^{\text{set}}(z)} + \frac{C(z)[1 - D'_L(z)]}{G_pH(z)KC^{\text{set}}(z)} \quad (21.28)$$

$D_s(z)$ is the same for both loops.

Cox and Shunta³⁰ compare the tracking dual algorithm $D'_L(z)$ and $D_S(z)$ with the conventional algorithms $D_L(z)$ and $D_S(z)$ in Figures 21.8 and 21.9 for the 20-tray distillation column whose transfer functions are given in Table 21.1. $D_L(z)$ and $D_S(z)$ are given by equations (21.7) and (21.9). $D'_L(z)$ is solved by substituting the appropriate terms in Table 21.1 into equation (21.26). K has a value of -730 to make $D'_L(z)$ physically realizable for a step-load disturbance. $C(z)$ was specified as in equation (21.6) for a feed-composition disturbance. $D'_L(z)$ alternatively could have been determined by equation (21.27).

$$D'_L(z) = \frac{0.637 z^{-1}}{1 - 0.368 z^{-1}} \quad (21.29)$$

Figures 21.10 and 21.11 show the closed-loop responses to a set-point change in x_2 from 0.057 to 0.01 for the conventional and tracking dual algorithms. The sampling period is 1 minute. The manipulative variable M (vapor boilup) is limited by an override to a maximum change of 11 moles/min while the controller output CO (expressed as moles/min vapor boilup) initially calls for a greater value. The conventional algorithm saturates and takes almost an hour to unwind (Figure 21.10). The tracking algorithm does not saturate and resumes control quickly. Figures 21.12 and 21.13 show the closed-loop responses to a 20 percent step increase in feed composition. Again control is improved with the tracking design.

Tracking PI Algorithm

The tracking structure is not unique to the dual algorithms. This same technique can be applied to restructure a discrete PI algorithm so that it does not saturate. Figure 21.14 compares the conventional loop and the restructured tracking form. A discrete version of a PI controller is found by approximating the continuous controller by rectangular integration³² (also known as "implicit Euler integration").

$$D(z) = \frac{K_c [\tau_R (1 - z^{-1}) + T]}{\tau_R (1 - z^{-1})} \quad (21.30)$$

K_c is the controller gain and τ_R is the reset time. $D'(z)$ is calculated from $D(z)$ and equation (21.27).

$$\begin{aligned} D'(z) &= 1 - \frac{K}{D(z)} = \frac{1 - K\tau_R (1 - z^{-1})}{K_c [\tau_R (1 - z^{-1}) + T]} \\ &= \frac{(K_c\tau_R + K_cT - K\tau_R) + \tau_R (K - K_c) z^{-1}}{K_c (T + \tau_R) \left(1 - \frac{\tau_R z^{-1}}{\tau_R + T}\right)} \end{aligned} \quad (21.31)$$

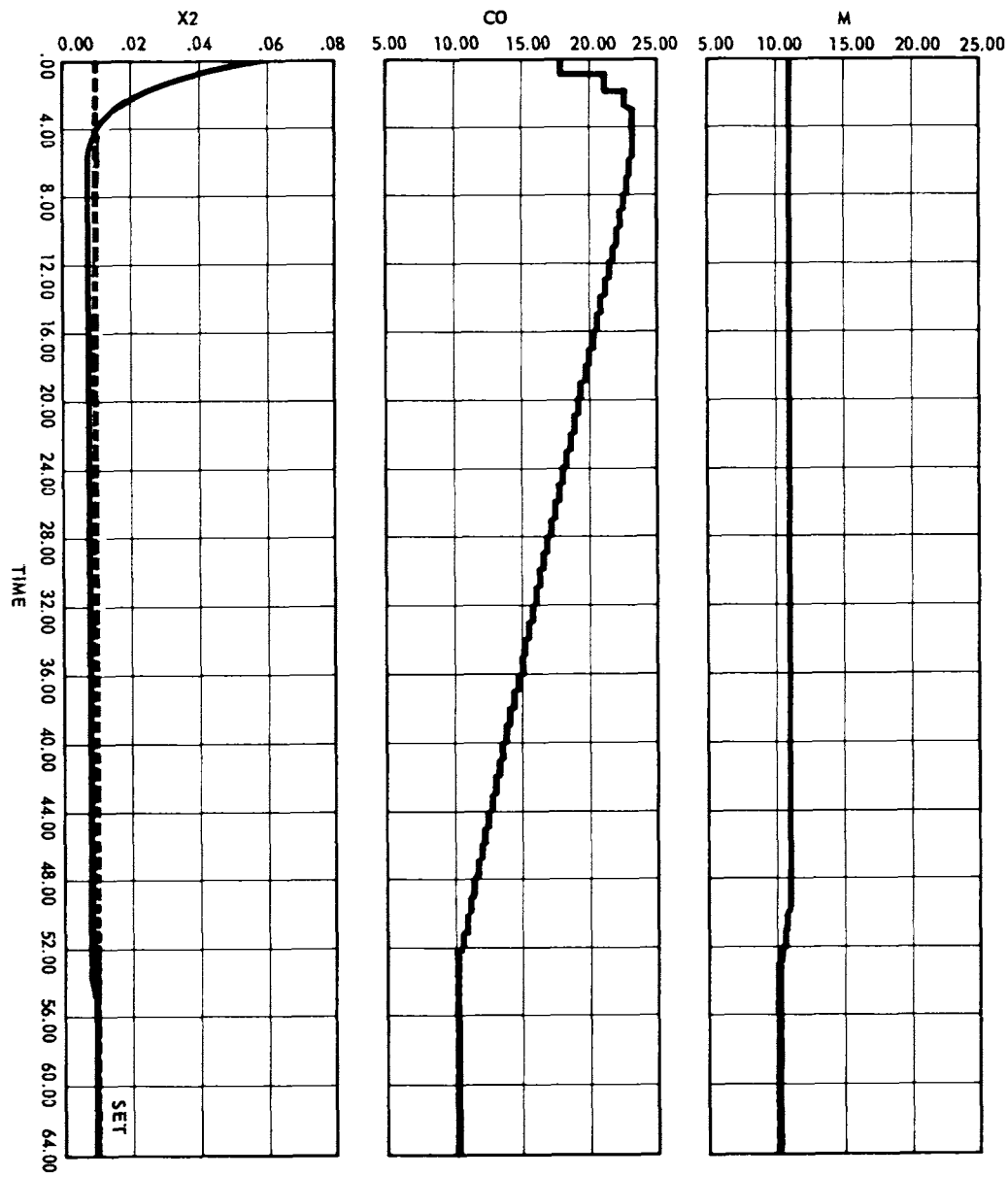


FIGURE 21.10
Conventional control of X_2 with setpoint disturbance

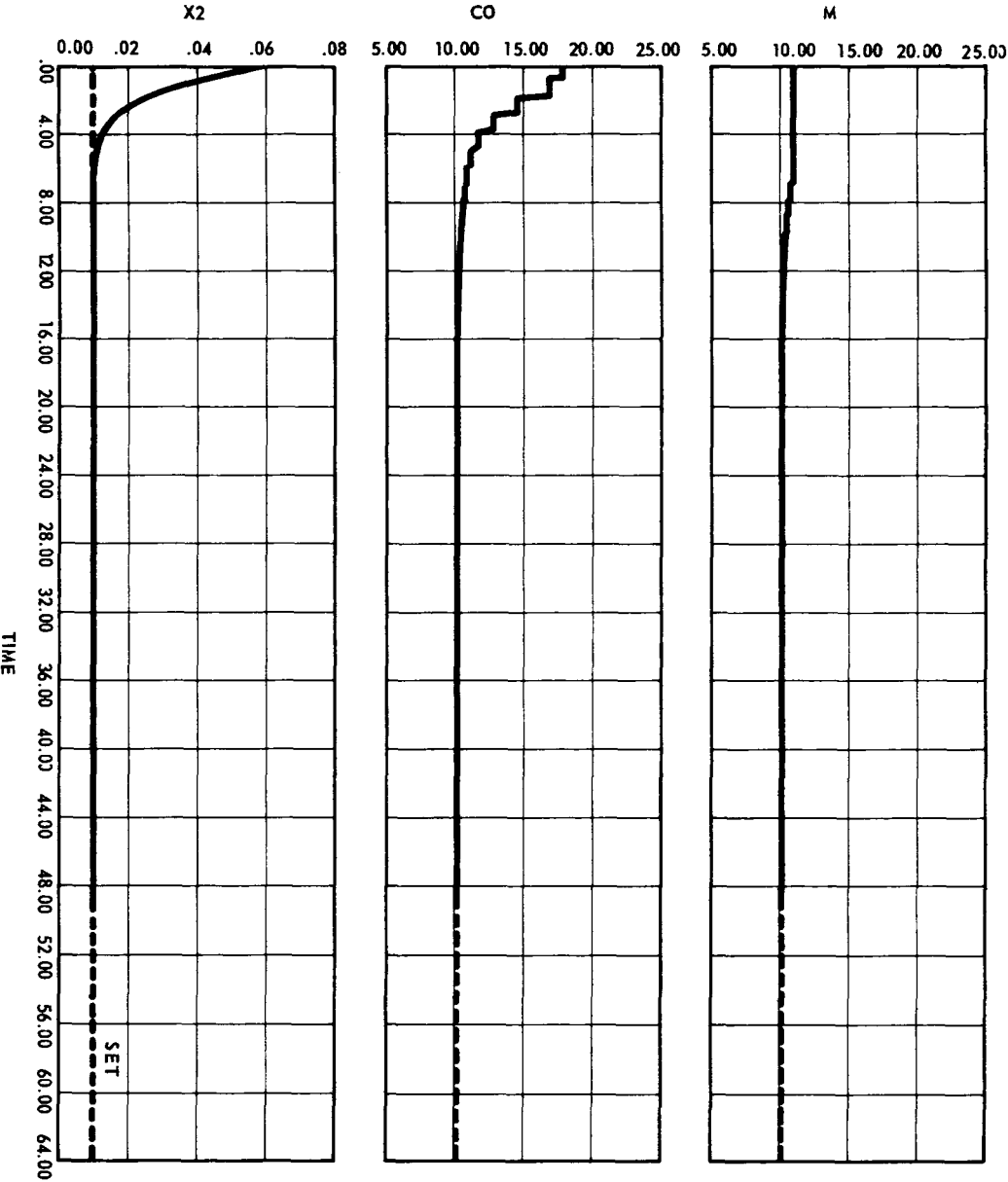


FIGURE 21.11
Tracking sampled-data control of X_2 with setpoint disturbance

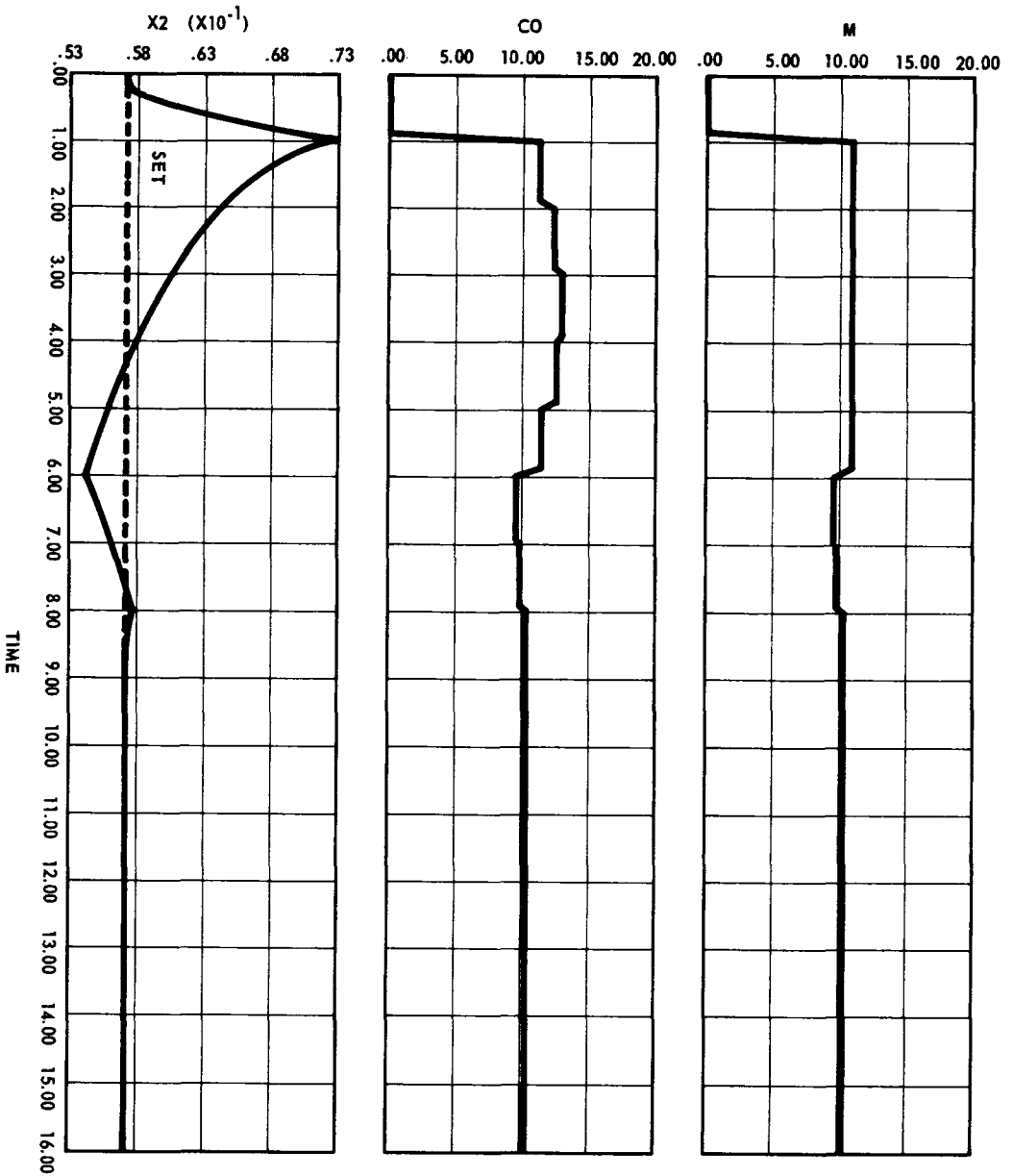


FIGURE 21.12
Conventional control of X_2 with feed composition disturbance

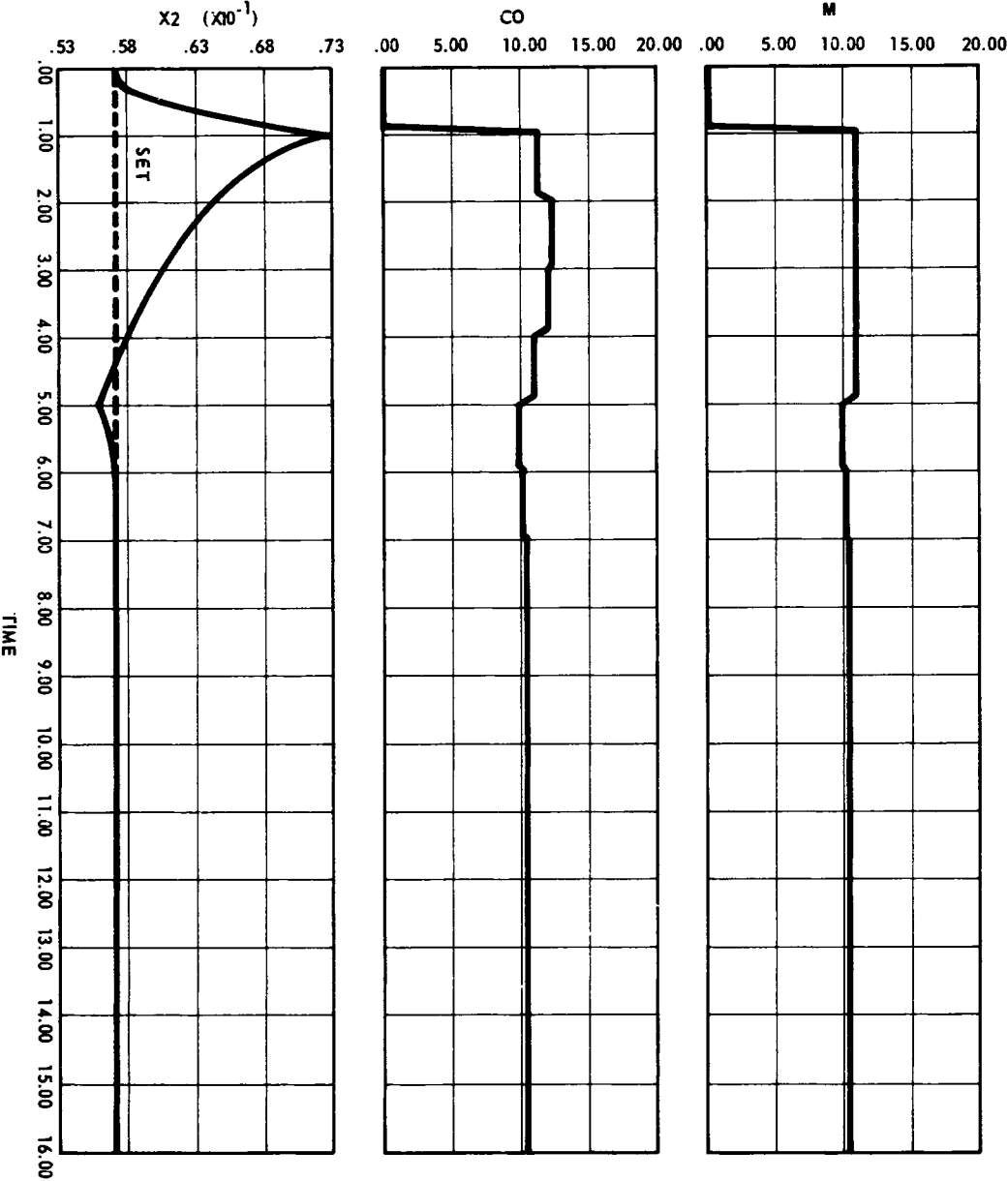


FIGURE 21.13
Tracking sampled-data control of X_2 with feed composition disturbance

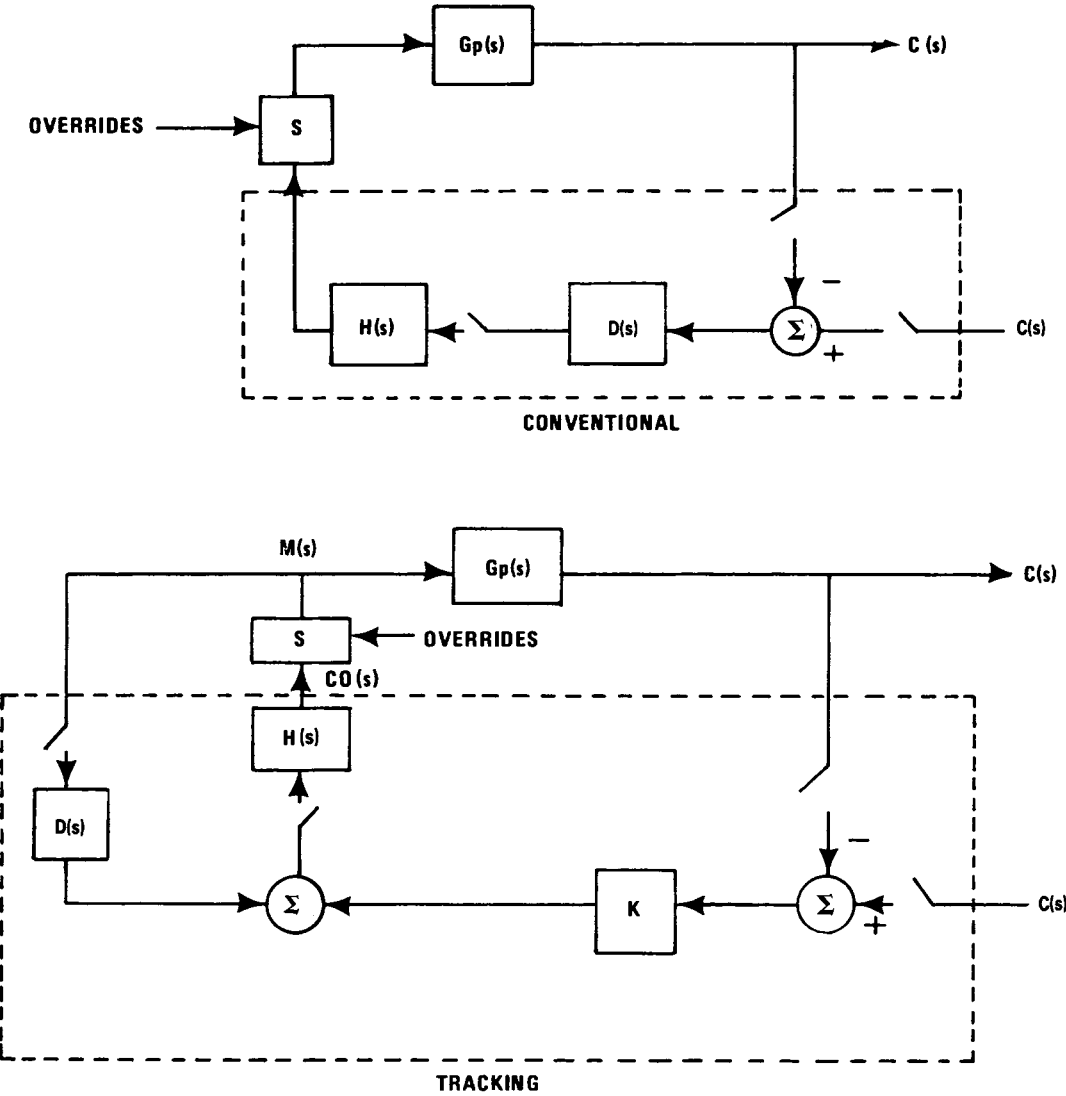


FIGURE 21.14
Comparison of conventional and tracking PI control

K is specified to make $D'(z)$ physically realizable by setting the left side of the numerator equal to zero.

$$K = \frac{K_c (\tau_R + T)}{\tau_R} \quad (21.32)$$

The final form of $D'(z)$ is found by substituting equation (21.32) into (21.31).

$$D'(z) = \frac{Tz^{-1}}{T + \tau_R - \tau_R z^{-1}} \quad (21.33)$$

The time-domain output of $D'(z)$ is a function of the past values of $D'(t)$ and the manipulative variable.

$$D'(t) = \left(\frac{\tau_R}{T + \tau_R} \right) D'(t - T) + \left(\frac{T}{T + \tau_R} \right) M(t - T) \quad (21.34)$$

The complete controller output is:

$$CO(t) = \frac{K_c (T + \tau_R)}{\tau_R} [C^{\text{set}}(t) - C(t)] + D'(t) \quad (21.35)$$

Therefore, the tracking algorithm prevents saturation when an override occurs, or almost any other break in the control loop, for that matter. The antisaturation capability is designed into the algorithm and requires no additional logic that takes up space and time in the computer. Khandheria³³ found that the feedback signal needs to be free of noise, and the A/D and D/A signal converters must be zeroed properly for best performance.

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